[COMMITTEE PRINT]

MITIGATING ACID RAIN WITH TECHNOLOGY: AVOIDING THE SCRUBBING-SWITCHING DILEMMA

REPORT

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AGRICULTURE RESEARCH AND ENVIRONMENT

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June 23, 1983

LETTER OF REQUEST

Gilbert Gude, Director
Congressional Research Service
Washington, D.C. 20540

Dear Gil:

We are writing to request that the Congressional Research Service compile a reader on emerging technologies to control acid rain precursor emissions (SO_2 and NO_x) from stationary sources. The current discussion on acid rain control is constrained by its focus on current technology, particularly flue gas desulfurization (FGD). By broadening the debate to include emerging technologies, it is possible that some seemingly intractable problems, such as high consumer costs and unemployment of coal miners, may become amenable to solution. However, for this to occur, the Congress must have an objective understanding of the status of these new technologies. We hope the reader which CRS will prepare for Committee publication will perform this service as well as assist the Committee in reviewing U.S. R&D priorities.

Dr. Robert Palmer of the Committee staff has contacted Mr. Robert E. Trumbule, who has indicated that he, along with Dr. Larry B. Parker of his section, would be willing to undertake this task. Staff contacts for this request will be Dr. Palmer at the Subcommittee on Natural Resources, Agriculture Research and Environment and Ms. Nancy Jeffrey at the Subcommittee on Energy Development and Applications. We look forward to working with these analysts and producing a useful and timely reference document.

Sincerely,

DON FUQUA
Member of Congress

JAMES H. SCHEUER
Member of Congress

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LETTER OF SUBMITTAL

July 20, 1983

Honorable Don Fuqua
Chairman, Committee on Science
and Technology
House of Representatives
Washington, D.C. 20515

and

Honorable James H. Scheuer
Chairman, Subcommittee on Natural
Resources, Agriculture Research and
Environment, Committee on Science and
Technology
House of Representatives
Washington, D.C. 20515

Dear Gentlemen:

I am submitting herewith a report entitled "Mitigating Acid Rain with Technology: Avoiding the Scubbing-Switching Dilema" which was prepared at your request. The report describes a number of technologies which either are available now or are emerging and will be available in the future to provide control of acid precipitation precusor pollutants—sulfur oxides and nitrogen oxides. As requested, the report further offers a structure for considering means by which development and use of these technologies might be encouraged.

A second part of this report contains selected documents from the literature which generally provide an overview of the development status of pollution control technology for oxides of both sulfur and nitrogen. Review of these documents demonstrates that several technologies for control of these pollutants are available and that many others are emerging as candidates for use in the next ten years.

The report was prepared under the direction of Robert E. Trumbule, Specialist in Energy Policy in our Environment and Natural Resources Policy Division. Mr. Trumbule and Larry B. Parker, Analyst Energy Policy wrote the CRS analysis with assistance from James W. Bourne, contractor. Mr. Bourne and Mr. Parker assisted in collecting, selecting and summarizing the documents used in the "Selected Readings" portion of the report.

Typing and editorial assistance were provided by the entire Environment and Natural Resources Policy Division production support staff, headed by Ms. Sandra Johnson.

Sincerely,

Gilbert Gude

Gilbert Gude
Director
Congressional Research Services

Enclosure

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MITIGATING ACID RAIN WITH TECHNOLOGY: AVOIDING THE SCRUBBING-SWITCHING DILEMMA

by

Larry B. Parker
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Revised
July 28, 1983

(XV)

MITIGATING ACID RAIN WITH TECHNOLOGY: AVOIDING THE SCRUBBING-SWITCHING DILEMMA*

I. INTRODUCTION

In discussing strategies to mitigate acid rain , the debate has generally revolved around two alternatives for reducing sulfur dioxide (SO2) and nitrogen oxide (NOx) emissions: (1) fuel switching facilities to lower sulfur coal, and (2) installing flue gas desulfurization (FGD) units on power plants. These alternatives present a dilemma for Congress between the capital intensiveness of FGD which can translate into higher costs to consumers (at least in the short-term); and, the coal industry effects of massive switching of facilities to low-sulfur coal, which would result in miner unemployment in high-sulfur coal regions.

This paradox regarding acid rain abatement is partially the result of a narrow view of potentially available technologies for controlling SO2 and NOx emissions from stationary sources. Technology are available now or will be in a very few years which can significantly reduce SO2 and/or NOx emissions without significantly reducing the efficiency of a plant or creating dislocations in the coal market. These technologies range from incremental combustion modifications for reducing NOx emissions to advanced combustion designs for efficient and clean burning of high-sulfur coal. Incentives could accelerate the use of these acid rain mitigating technologies, if such a course is considered appropriate.

The purpose of this paper is two-fold: first, to provide an overview to the accompanying articles on the advancement of SO2 and NOx control technologies, and secondly, to discuss factors and incentives for influencing their widespread commercial use in conjunction with an acid rain mitigation program. In this vein, a considerable amount of generalization must be done. Readers interested in the specifics of any one process are urged to seek out the articles contained within on those technologies. This overview attempts to give the reader a feel for the current situation regarding SO2 and NOx technologies and therefore should be considered illustrative of current efforts, not comprehensive. The paper does not address the underlying controversy surrounding the merits of an acid rain program.

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*The authors would like to acknowledge the research contribution of Mr. James W. Bourne, contractor, in the preparation of this summary.

The paper is divided into five parts: (1) background on the categories of technologies currently available or anticipated in the future, (2) examples of the current research activity in the area and cost estimates for various technologies, (3) identification of factors affecting the introduction of technologies on a commerical scale, (4) analysis of various incentives and strategies for accelerating the introduction of selected technologies as a part of a proposed acid rain mitigation program, and (5) a conclusion.

II. BACKGROUND

Control of SO2 and NOx emissions can be done either before, during, or after combustion. Pre-combustion strategies include cleaning the fuel before combustion or using low-sulfur fuel. Combustion controls include modifications to the burners or the addition of a reagent to react with targeted pollutants. Post-combustion processes, the most commonly employed pollution control method on new power plants, currently consist of some form of flue gas desulfurization (FGD) which reduce SO2 emissions after fuel combustion. Finally, advanced systems are currently under development that burn fossil fuels differently than conventional processes, emitting lower levels of SO2 and NOx.

Pre-Combustion Controls

Sulfur oxide emissions result from the sulfur content of the fuel, while NOx emissions occur both from the nitrogen in the fuel ("fuel-NOx") and from the high temperatures in the combustion chamber ("thermal NOx"). The sulfur in the coal is found in two forms: inorganically bound, often with iron as pyrites, and organically bound. Although the latter is difficult to remove, the inorganically bound sulfur can be removed by physical means such as crushing and washing.

Physical coal cleaning is a proven pre-combustion control technology. Coal has a lower specific gravity (usually 1.3-1.75) than the impurities associated with coal (such as sulfur) which generally have a specific gravity greater than two. Hence, placing the coal in a solution with a specific gravity between the two causes the coal to float and the impurities to sink. The coal is then removed and dried. Various physical processes are described in Table 1.

Table 1: Characteristics of major wet-coal cleaning devices

Device	Medium	Principle of separation	Sizes typically handled	Unit capacities	Specific gravity range	Cleaning efficiency	Comments
Coarse coal jig	Water	Stratification	6 × ¼ in.	High	1.60-1.90	Good	Typically used to clean coarse coal at relatively high specific gravities. Efficiency decreases rapidly at sizes below ½ in.
Fine coal jig	Water	Stratification	½ in. × 28M	High	1.60-1.90	Good	Reported to be as efficient as concentrating tables when handling compatible size. High unit capacity.
Hydrocyclone	Water	Centrifugal forces in ascending vortex	½ in. × 100M	High	1.60-1.90	Poor	Poor cleaning efficiency, but can be used as 'scalpers' ahead of more efficient devices.
Concentrating table	Water	Stratification and hindered settling	⅜ in. × 200M	Low	1.50-1.80	Good	Provides good separation when near-gravity material is less than 10% of feed. Requires very low water-to-solids ratio.
Heavy-medium vessel	Fine magnetite in suspension	Float and sink	6 × ¼ in.	High	1.30-1.80	Excellent	Efficiency of separation near theoretical, even when feed contains large amounts of near-gravity material.
Heavy-medium cyclone	Fine magnetite in suspension	Float and sink plus centrifugal forces	½ in. × 28M 1½ in. × 0	High High	1.30-1.80 1.30-1.80	Excellent Very good	Excellent separating efficiency over wide range of specific gravities. Raw feed must be deslimed. Can replace multiple cleaning circuits. Requires no raw-coal desliming.
Flotation	Water	Particle attachment	28M × 0 100M × 0	High Low	N/A N/A	N/A N/A	Not effective for oxidized coal or for removal of pyritic sulfur. More expensive than other cleaning methods.

Source: Coal Cleaning to Improve Boiler Performance. Power, September 1981.

The cleaned coal provides for more efficient burning, reduced fuel transportation costs, and lower sulfur dioxide levels emitted. Although the economics are site-specific, study has shown that often it is economical to clean coal prior to combustion, regardless of any environmental benefits. 1/ The major disadvantage of the system involves a relatively low level of sulfur removal. Because the process removes only pyritic sulfur, it generally allows only a 20-35 percent sulfur reduction. Also, there is a tradeoff between increasing sulfur removal and increasing the Btu losses associated with the cleaning process.

1/ Electric Power Research Institute. Impact of Coal Cleaning on the Cost of New Coal-Fired Power Generation. Report CS-1622. March 1981.

To reduce sulfur further before combustion, development of chemical coal cleaning processes are underway. Unlike physical cleaning, chemical processes also remove organically bound sulfur, in addition to a high percentage of pyritic sulfur. Chemical cleaning is a much more complicated process, requiring chemical reagents to break apart the bound sulfur. Although it is a more expensive process, a higher sulfur removal rate may reduce or eventually eliminate the need for further control modifications, thus making this approach more economically feasible.

Combustion Controls and Modifications

Controlling emissions during combustion is a second area where reductions are possible. Combustion control technologies currently under development are often less complicated than post-combustion control techniques, they produce waste products which are disposed of more easily, and the capital and annual costs are, in some cases, projected to be substantially less than current post combustion systems such as FGD and selective catalytic reduction (SCR) for the same reduction capability. For SO2, control generally consists of adding alkaline absorbents, such as limestone, to react with the sulfur-laden gas before the gas exits the combustion chamber. For NOx control, there is a variety of techniques for controlling the supply of air and fuel, along with flame temperatures in the combustion chamber, thus reducing fuel and thermal NOx.

Current research and development of combustion modifications have focused on systems that simultaneously control both SOx and NOx, such as Limestone Injection Multistage Burners (LIMB), and Atmospheric Fluidized Bed Combustion (AFBC). Table 2 summarizes some of the advantages and disadvantages of these two control technologies.

TABLE 2: Characteristics of Selected Simultaneous SOx and NOx Control Techniques

Technique	Current Status	Advantages	Disadvantages	Projected Costs
LIMB	Under development, retrofit operating in West Germany	Easily retrofitable. Inexpensive. 50-70% of SO2 and NOx reductions.	For higher SO2 removal use of another control technique will be necessary.	Capital costs \$27-\$37/Kw.* Annual costs 5.6-6.1 m/kwh.
AFBC	Commercial facility by late 1980s 2/	Can burn high sulfur coal with good emission control. Dry waste can be disposed of more easily.	Potential problems with bed stabilization at larger sizes. Presence of quick lime in waste may present disposal problem.	Lower costs than current combustion technologies.**

* EPA estimates in 1982 dollars.
** TVA's current projection

Source: Data compiled by CRS from articles attached.

Of these technologies, LIMB is a particularly promising control process. It has the potential to reduce both SO2 and NOx emissions by 50 to 70 percent, and is designed to be retrofitted to existing facilities. Furthermore, LIMB is projected to be inexpensive to install and operate. The capital costs of LIMB are projected to be between one-seventh and one-tenth that of FGD. 2/

Another combustion system which has received substantial attention, especially from the Tennessee Valley Authority (TVA) is Atmospheric Fluidized Bed Combustion (AFBC). In this process, powdered coal and limestone is injected onto a "fluidized-bed" raised by air pressure from below. Sulfur oxides react with the limestone, while nitrogen oxide formation is reduced due to lower combustion temperatures. Current research suggests that AFBC will provide

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2/ U.S. Environmental Protection Agency. LIMB/LOW NOx Burners - EPA's Program to Develop Low Cost SOx/NOx Controls for Coal-fired Boilers. Revised March 1983.

substantial SOx and NOx reductions at costs less than that for conventional systems. 3/

For control of NOx alone, a number of modifications in combustion techniques have been developed, some of which lead to improved combustion efficiencies. Dual register burners, staged combustion, use of overfire air, reduction of excess air and flue gas recirculation all have potential for reducing NOx emissions. Table 3 summarizes some of the advantages and disadvantages of some of these modifications.

Post-Combustion Controls

Controlling emissions after the combustion phase has been primary emphasis of U.S. SO2 control efforts. Wet scrubbers using lime or limestone are the major means of active SO2 control currently used in the U.S. today, providing over 90 percent SO2 removal capability. This focus on a technological response to SO2 emissions is almost unique among the industrialized countries which tend to favor switching plants to lower-sulfur coals to control emissions. Indeed, when permitted, U.S. utilities have traditionally preferred to fuel switch rather than install FGD units. This reluctance is the result of cost and problems such as plugging, scaling and other troubles that plagued early, first-generation wet-scrubbers; problems which have been somewhat rectified in recent years.

With the Clean Air Act mandating a percentage reduction in SO2 emissions, efforts began to develop a cost-effective dry scrubbing system as an alternative to wet FGD. Dry scrubbing offers some distinct advantages in the control of SO2. Lower energy and less manpower requirements, as well as greater dependability and simplicity may make dry processes more cost effective

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3/ Tennessee Valley Authority. Atmospheric Fluidized Bed Combustion: Technology Development and Demonstration. November 1, 1982.

TABLE 3: Characteristics of Selected Combustion Modifications for NOx Control

Modification	Current Status	Advantages	Disadvantages	Costs
Low excess-air combination	Currently available	Process is more efficient way to burn fuel. Results in 17-38% NOx reductions for all fuel types.		Inexpensive way to reduce NOx levels. Can save money by more efficient fuel burning. Costs in the tenths of mills per kwh.*
Flue gas recircultion	Currently available	Reduces NOx up to 60% for certain fuels. Also reduced soot levels.	Restricted to low nitrogen fuels since this process reduces thermal NOx only. Decrease of optimum flame temperatures causes lower fuel efficiency.	Installation costs are greater than other low NOx burner systems.
Staged combustion	Currently available	NOx reductions of 30-50%, retrofitable in some cases.	Potential problem with unburned fuel, high level of corrosion and high maintenance requirements.	Low-cost way to reduce NOx emissions. Cost in the tenths of mills per kwh.*
Dual Register Burners	Currently available	In use since 1972. NOx reductions up to 60% possible using oil or coal.		Capital costs are estimated at $5/kw. Annual operating costs of 0.03 mills/kwh.**

* CRS estimate in 1980 dollars
** EPRI estimate in 1980 dollars

Source: Data compiled by CRS from articles attached.

than wet systems to operate, especially for sources that burn lower sulfur coal and do not need to meet a 90 percent reduction requirement. 4/

The drawback of dry systems include a low SO2 control level and the inability to work effectively on high sulfur coals (more than three percent). Dry scrubbing reduces SO2 emissions by only 70%, versus 90% reductions from wet systems. Also, the cost advantage of the technology versus wet FGD decreases as the sulfur content of the coal increases toward the technology's current three percent maximum. This is because of the quantity of relatively high-cost reagent necessary to react with the sulfur becomes too costly for the process to compete with wet FGD. 5/ Table 4 summarizes the general advantages and disadvantages of wet and dry FGD systems.

A general advantage of post combustion control techniques is that they may be combined with pre-combustion controls, such as coal preparation, to increase the overall reduction of SO2. In particular, the combination of physical coal cleaning and wet or dry scrubbing of only a portion of the flue gas may be economically attractive. Also, the combination can increase the reliability of the control systems. 6/

Post-combustion controls have also been developed for nitrogen oxides. Two such processes include selective catalytic reduction (SCR) and selective non-catalytic reduction (SNR). SCR can provide up to 95 percent NOx control; SNR up to 60-80 percent. SCR uses a catalyst, such as copper oxide, to assist a reaction between SO2 and ammonia (NH3) to form nitrogen

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4/ Satriana, M. (ed.) New Developments in Flue Gas Desulfurization Technology. Noyes Data Corporation, Park Ridge, N.J., 1981. p. 185.

5/ Ibid.

6/ U.S. Department of Interior, U.S. Environmental Protection Agency. Coal Cleaning with Scrubbing for Sulfur Control: An Engineering/Economic Summary. EPA 600/9-77-017, August 1977.

TABLE 4: Characteristics of Wet And Dry FGD Systems

Wet Processes	Energy Requirements	Advantages	Disadvantages	Costs
Lime/Limestone Sodium Carbonate Welman Lord Double Alkali Others	Must reheat flue gases before emissions.	More verstile systems for different coals and sulfur contents. High SOx removal (up to 95%). Some particulate removal. Greatest operating experience.	Large sludge production in some cases. Need to control pH level. Plugging, scaling, and corrosion. Large water requirements.	Require use of more expensive alloys for construction. More manpower needed. Least expensive systems currently in use.
Dry Processes	Require 25-50% less energy than wet systems.	No problems of scaling or plugging. Less manpower needed to operate dry systems. No need to control pH level carefully. Higher availability with simpler system (no sludge control mechanisms) Uses much less water, and of lower quality. Lower maintenance requirements.	Cannot usually use dry systems on coals with sulfur contents above 3%. More expensive reagents required.	Can use carbon steel for vessels. Reagents more expensive than with wet systems. Use less energy--no reheat of flue gas Less equipment needed since sludge processing not involved. Cost effective way to reduce SO2 in some cases.

Source: Data compiled by CRS from articles attached.

and water. Unfortunately, in some cases, the additional formation of ammonium sulfate and ammonium bisulfate can lead to equipment plugging and clogging. SNR is similar to SCR, except that it does not use a catalyst, but depends on the high temperature of the gases to drive the reaction with ammonia.

Generally, SCR costs about two to three times as much SNR in terms of annualized costs. Both technologies are commercially available with SCR used currently in Japan and SNR on oil refinery preheaters in California. 7/

A final area of work on post-combustion control is simultaneous SOx and NOx control technologies. The goal is to develop single process systems for what is now accomplished with two systems, such as FGD for SO2 and SCR for NOx. As with FGD, there are both wet and dry systems. Wet simultaneous control technologies are currently thought to be economically uncompetitive compared to separate SO2 and NOx control technologies, 8/ although one wet process (Sulf-X) has been reported to be economically attractive. This wet process is regenerative, producing a saleable by-product sulfur. 9/

Dry processes for simultaneous control have been given greater emphasis, due to their potential lower overall costs for NOx and SOx control. 10/ There are several systems available, including the Shell copper oxide process and the carbon absorption process. These control strategies should have high (90% or better) SOx reductions and upwards of 70% NOx removal. Demand for either of these systems will not be great unless NOx removal

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7/ Parker, Larry B. and Robert E. Trumbule. Opportunities for Increased Control of Nitrogen Oxides Emissions From Stationary Sources: Implications for Mitigating Acid Rain. CRS report 82-217, December 9, 1982.

8/ Dalton, S.M. Current Status of Dry NOx-SOx Emission Control Processes. Presented to 1982 Joint Symposium on Stationary NOx Control. Dallas, Texas November 1-4, 1982

9/ Ellison, William, and Edward Shapiro. The Sulf-X Flue Gas Desulfurization Process. Presented to Tenth Energy Technology Conference and Exposition, Washington D.C. February 28, 1983.

10/ Dalton, S.M. Current Status of Dry NOx-SOx Emission Control Processes.

requirements become stricter, due to their higher costs in comparison to current systems capable of meeting current emission standards. Table 5 compares the advantages and disadvantages of two of these dry simultaneous control methods.

D. Advanced Technologies

Processes included in this section are more complex than the combustion modification or flue gas treatment processes used currently. Some advanced technologies involve refinements of combustion processes, while others employ revolutionary techniques for the production of energy.

TABLE 5: Characteristics of Two Dry Simultaneous SO2 and NOx Control Technologies

Technique	Current Status	Advantages	Disadvantages	Costs
Copper Oxide	Pilot plant testing in U.S. 40 Mw commercial plant operating in Japan since 1973.	90% SO2 removal 60-70% NOx removal. Low waste by-product level due to regenerative system. No plugging of reactor due to low formation of ammonium sulfates.	High temperatures in process make retrofitting difficult. Process requires about 5% of energy produced from boiler.	High capital and annual operating costs. Capital costs $238 per Kw. Annual operating costs 5.6 mills/kwh.*
Carbon Absorption	Commercial plant now under construction in West Germany (SO2 control only) NOx control still at pilot test level. Prototype testing in U.S. and Japan.	SO2 control as high as 98% 75% NOx reductions. Waste produce reduced by regeneration saleable by-product produced.	Difficulty in controlling flue gas temperatures. (Increased temperatures reduce SO2 control, but slightly increase NOx control) Sorbent deterioration with use.	Expensive in comparison to current FGD systems. Little demand unless strict NOx emissions required.

* Updated EPA estimate.

Source: Data compiled by CRS from articles attached.

Another reason for their separate treatment here is their state of development. These systems are still several years away from commercial use, but have the potential to burn fossil fuels more efficiently and emit lower levels of SO2 and NOx than systems currently in use. These systems generally are not retrofittable, but may have wide future application in burning coal in an environmentally-sound manner. However, it should be noted that more research and development will be necessary before commercialization of these technologies is contemplated.

Electron-beam irradiation is one advanced system which is progressing well through developmental stages. Several pilot-plants have operated in the U.S. and Japan, and the producer is now attempting to secure support for a demonstration facility in the U.S. The process involves the dissociation of NOx and SOx by electron beams to form free radicals, and subsequent reactions with ammonia to form ammonium sulfate and ammonium bisulfate. Two process variations are being developed, one with a regenerable by-product and the other with a throwaway system. The process is projected to be approximately 25 percent less expensive than a combined use of wet flue gas scrubbing and selective catalytic reduction to reduce SOx and NOx. 11/

Pressurized fludized bed combustion (PFBC) involves a fluidized-bed burning process, but operates at higher pressures than does atmospheric pressurized fluidized-bed combustion. Since the reaction occurs in a smaller space, the unit is smaller, which may reduce capital and operating costs. PFBC produces less waste than AFBC, and is less sensitive to clogging and plugging of the injection system. PFBC is more efficient, controls NOx better than AFBC and uses less limestone for the reaction with SOx. 12/

11/ Dalton, S.M. Current Status of Dry NOx-SOx Emission Control Process.

12/ National Academy of Sciences. Assessment of Advanced Technology for Direct Combustion of Coal. National Academy of Sciences. Washington, D.C. 1977.

A second emerging technology for converting coal into electricity is the gasification-combined cycle (GCC) power plant. A GCC plant operates by first gasifiying the coal and then burning the resulting clean gas in a combustion turbine at temperatures above 2000 degrees Fahrenheit. The exit gas is transmitted to a heat-recovery steam generator which uses it to power a steam turbine.

GCC is potentially retrofittable, thermally more efficient and most important, is projected to be cost-effective compared with conventional generation systems. A potentially important base load process, GCC is also very flexible in terms of energy output. The system can easily fluctuate output to meet energy demand. 13/

Combustion of fossil fuels with magnetohydrodynamics (MHD) could virtually eliminate SO2 emissions and substantially reduce NOx emissions. In this process, electricity is generated directly from the thermal energy, bypassing the transfer of heat to mechanical energy. An electrically conducting gas is channeled through a duct at high speed in the presence of a magnetic field. This process creates a voltage drop through the stream of gas.

In addition to low pollution emissions, MHD systems are much more energy efficient than conventional processes. While current electric utility plants operate at 33-40 percent efficiencies, MHD systems could operate at efficiencies around 50 percent, resulting in significant fuel savings. 14/

13/ Quentin, George. Coupling a Gasifier to a Combined-Cycle Plant. EPRI Journal, April 1981. p. 22-27.

14/ Bienstock, D. and V. M. Maslennikov. Protection of the Biosphere. In Petrick, M. and Shumyatsky, B. Ya. (eds.) Open-Cycle Magnetohydrodynamic Electric Power Generation. A Joint USA/USSR Publication by Argonne National Laboratory, 1978.

III. CURRENT RESEARCH AND COST PROJECTIONS

A. Examples of Current Activity

Research is currently underway involving numerous processes to reduce SO2 and NOx emissions from stationary sources more efficiently and reliably than current systems. A variety of processes is emerging to meet specific emission source characteristics, such as furnace type, fuel type, sulfur content and other variables. These emerging technologies are in various stages of development. Table 6 reviews the availability of some of these technologies.

These technologies are the result of Federal, private and combined federal-private efforts. An example of Federal research is the development of LIMB technology. For several years, the EPA actively supported the development of LIMB technology. It is largely due to EPA that LIMB has advanced to its current state. During the Reagan Administration, however, funds for continued development of LIMB by the EPA have been reduced drastically, as has funding for a demonstration effort. This action, some argue, has lengthened the time required for this technology to reach commercialization.

Another example of Government activity is TVA, which is actively pursuing development of AFBC. TVA has proposed building a 160 mw plant that will burn high sulfur, Illinois coal. The AFBC facility is projected to be on line by 1989, and be less expensive than other combustion technologies. The U. S. Department of Energy has also sponsored pilot plants for various control technologies, including the Wormser grate. The Wormser grate is an industrial-sized fluidized-bed combustion device; two commercial facilities using the grate have been built. One system provides steam for a meat-packing plant in Amarillo, Texas and the second was retrofitted to a boiler at a university in Massachusetts. The Department of Energy has also agreed to fund the expansion of a Sulf-X demonstration plant which began operations in 1982.

TABLE 6: Current Status of Selected Control Technologies.

	TECHNOLOGY	COMMERCIAL STATUS*
1.	Physical Coal Cleaning	Available
2.	Chemical Coal Cleaning	Under Development
3.	Wet FGD	Various Processes Available
4.	Dry FGD	Various Processes Available
5.	Low NOx Burners	Various Designs Available
6.	Selective Catalytic Reduction	Available (Extensive use in Japan, limited use in U.S.)
7.	Selective Non-Catalytic Reduction	Available
8.	LIMB	Demonstration plant within 3-5 years
9.	MHD	Under development
10.	AFBC	Commercial plant by late 1980s
11.	PFBC	Commercial plant by mid-1990s
12.	GCC	Demonstration plant under construction Commercial plant by 1990s
13.	Electron Beam	Under Development
14.	Carbon Absorption	Pilot plant in U.S. (mid-70s) Demonstration plant in Japan soon
15.	Copper Oxide	Available

* Research is a continuing process with refinements being made and variations on designs being tested. Hence, "available" only implies that a general design is commercially available for use.

Source: Data compiled by CRS.

The Electric Power Research Institute (EPRI) has also been responsible for considerable development of control technologies for utility plants. As the research arm of the utility sector, EPRI has financed numerous research and demonstration projects in order to increase the quantity and quality of control devices. EPRI, along with Bechtel, General Electric, Southern California Edison Co., and Texaco, have pioneered the development of gasification-combined cycle by subsidizing a plant for Southern California Edison, which should start operations in early 1984.

Other examples of industry research and development can be found with the major oil companies. For example, Exxon has developed an SNR device, called thermal DeNOx, for significantly reducing NOx emissions. The system has been developed on oil and gas fired steam boilers and utility boilers, as well as coal fired boilers. Shell Oil has developed the copper oxide process to simultaneously reduce SOx and NOx emissions.

Boiler manufacturers, such as Babcock and Wilcox, Foster and Wheeler, Combustion Engineering, and Riley Stoker are also working on technologies to control emissions, particularly NOx. Examples of their work include development of dual register burners, and Combustion Engineering's Low-NOx Concentric Firing System (LNCFS). 15/

Finally, there are numerous other groups actively researching and refining emission control technologies. The air pollution control industry has become a significant industry in the U.S. with an estimated 111,000 man-years' worth of employment involving fabrication, design, operation,

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15/ Parker, Larry B. and Robert E. Trumbule. Opportunities for Increased Control of Nitrogen Oxides Emissions From Stationary Sources: Implications for Mitigating Acid Rain. CRS Report 82-217 ENR, December 9, 1983.

maintenance and repair, and other jobs. 16/

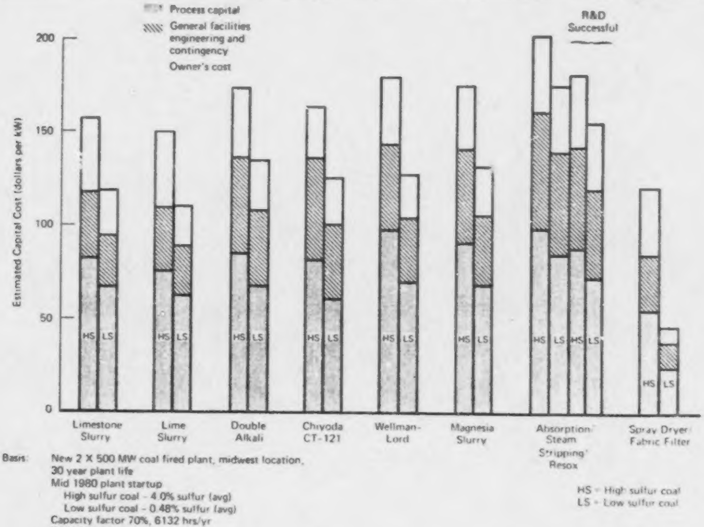
B. Costs

Industries that would be required to reduce emissions of SOx and NOx emissions under an acid rain program have voiced their strong concern that reductions would prove very expensive. Industry leaders feel that reductions to lessen the effects of acid precipitation should not be implemented until there is conclusive scientific evidence to affirm a connection between utility and industry emissions and acid rain.

Current standards mandate a percentage reduction in SO2 emissions at new coal burning utility facilities. To meet this standard, wet FGD throwaway systems have been employed widely, and have logged the most operating experience of any system. The capital cost for installing throwaway FGD systems into new facilities adds, on average, about 10% to the cost of construction. The cost to retrofit facilities currently in operation are generally one-third more than for new facilities. Figure 1 presents the estimated cost of various systems in dollars per kilowatt (kw). The first four processes are wet throwaway systems, which have waste disposal problems. Plugging, scaling and corrosion have also affected the performance dependability for these wet systems. The next three systems presented in the figure are regenerative processes. These systems can produce a saleable by-product, or one that is more easily disposable. The reagents used are often recyclable too, which will often reduce operating expenses. The final system, the spray dryer, is a dry process.

16/ H & W Management Science Consultants. Employment in the Air Pollution Control Industry. Prepared for the Industrial Gas Cleaning Institute, February 1982.

FIGURE 1: Flue Gas Desulfurization Estimated Capital Cost



Basis: New 2 X 500 MW coal fired plant, midwest location,
30 year plant life
Mid 1980 plant startup
High sulfur coal - 4.0% sulfur (avg)
Low sulfur coal - 0.48% sulfur (avg)
Capacity factor 70%, 6132 hrs/yr
85% SO_2 removal

SOURCE: National Academy of Sciences. 1980

As noted, wet systems have dominated the market for the past decade, and have the most on-line experience. However, due to the problems of dependability and waste disposal experienced by wet scrubbers and despite considerable improvements in their operation, alternatives have been pursued through research and development. While specific cost projections are not available for many of the processes under development, Table 7 presents some representative retrofit costs for some categories of technologies examined in this reader. As indicated, in specific situations, all of the technologies listed could be cost effective in comparison with wet FGD (although not necessarily with fuel switching). This is particularly true for LIMB technology which is projected to have capital costs of between 10 and 15 percent that

TABLE 7. Retrofit Control Technologies for a Typical 300 MWe Eastern U.S. Coal Fired Utility Boiler

	Lime/Limestone FGD	Lime Spray Drying	Physical Coal Cleaning	Limestone Injection in Multistage Burners	Low-NOx Burners (NOx Control Only)
Technical Status	Commercial	Early Commercial	Commercial	Developmental	Early Commercial
SO2/NOx Removal Efficiency	90% SO2 Reduction	75% SO2 Reduction	20-30% SO2 Reduction	50-60% SO2 Reduction 50-60% NOx Reduction	60-70% NOx Reduction
Capital Investment ($/kWh)*	260	190	46	27-37	2.0-7.1
Annual Operating Expense (mills/kWh)*	6.4	5.7	1.7	3.1	0.05
Levelized Annual Revenue Requirement (mills/kWh)*	23	18	4.9	5.6-6.1	0.2-0.4

*/ Costs expressed in December 1982 dollars, assuming a remaining boiler life of 15 years and a 40% operating factor.

Source: U.S. EPA LIMB Handout, February, 1983 of FGD. This cost advantage for LIMB is enhanced further by factoring in the costs of NOx control not achieved by FGD. However, wet systems probably will continue to dominate the SO2 control market, due to industry's reluctance to use other systems with less operational experience, unless concerted efforts are made to encourage development and demonstration of these potentially more cost-effective technologies.

Significant reductions in NOx emissions, down to .4 or .3 lb. per million Btu, are possible through advanced combustion modifications. The annualized costs of such modifications are estimated in the range of tenths of mills per kilowatt-hour (constant 1980 dollars). If a stricter standard were desired and could not be met by combustion modifications in a timely fashion, then the addition of thermal DeNOx could, for example, be the next step. Indeed, thermal DeNOx, in addition to combustion modifications may be able to achieve a .2 lb. per million Btu standard. Such a combination is estimated to cost under 1 mill per kilowatt-hour (annualized cost, constant 1980 dollars). A very strict standard of .1 lb. per million Btu could be achieved either through selective catalytic reduction techniques or combined SOx/NOx reduction techniques. Such equipment would cost in the range of 2 mills per kilowatt-hour, which would be about half the cost of flue gas desulfurization (annualized cost, 1980 dollars.) 17/

For new combustion technologies, costs have been estimated for atmospheric fluidized bed combustion and gasification-combined cycle. As calculated by TVA (see Table 8), both AFBC and GCC have the potential to reduce SOx and

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17/ Parker, Larry B. and Robert E. Trumbule. Lowering NOx Emission Standards for Stationary Sources: Analysis of Current and Future Reduction Potential. CRS unnumbered report, June 21, 1983.

TABLE 8: Cost and Performance Data for Coal-Based Generation Technology

	CONVENTIONAL PULVERIZED COAL			AFBC	GCC	
	SUBCRITICAL 2400 PSI FGD	SUPER-CRITICAL 3500 PSI FGD	LOW HEAT RATE 4500 PSI/ CHIYODA FGD	AFBC	TEXACO 2000°F	TEXACO 2600 F
FIRST COMMERCIAL SERVICE	CURRENT	CURRENT	1992	1993	1991	2000
NET HEAT RATE (BTU/KWH)	9860	9450	8420	9640	9460	8465
NET EFFICIENCY (%)	34.6	36.1	40.6	35.3	36.1	40.3
$/KW (IN SERVICE JANUARY 1981)	995	1000	1105	820	1030	965
LEVELIZED MILLS/KWH (1981-2010)	85	81	78	77	81	79

NOTES: AFBC MEANS ATMOSPHERIC FLUIDIZED BED COMBUSTION.
GCC MEANS GASIFICATION COMBINED CYCLE.
FGD MEANS FLUE GAS DESULFURIZATION.

Source: Tennessee Valley Authority, Division of Energy Demonstrations and Technology (August 1982) and the Electric Power Research Institute 1981 Technical Assessment Guide.

NOx more economically than current systems using FGD. If standards for NOx are strengthen in the future, this cost advantage would increase since both emit NOx at levels below current standards.

IV. FACTORS INFLUENCING INTRODUCTION OF TECHNOLOGIES.

Many of the technologies being discussed here are beyond the initial research and development stages of the commercialization process. The basic research has been done and practical applications have been invented and designed. As noted, this activity is the result of Federal, private and combined Federal-private research and development efforts. Also, considerable foreign activity has broadened the base of research and development for these technologies.

However, this linkage between public and private research and development activities begins to break down somewhat during the demonstration stage. Fundamental questions regarding the proper extent of Federal research assistance to private industry begin to impede the relationship. These questions have come into sharp focus with the current Administration's policy that once the innovation has reached the "proof of concept" stage, it is the business of private industry to determine its marketability and proceed with further demonstration and commercialization activities. The purpose is to allow marketability to choose the most useful course of innovation, ensuring both usefulness and economy.

Hence, the decision on whether or not to extend incentives to private enterprise to accelerate commercialization of a technology depends on one's assessment of private industry's current and future reaction to the technology. While private industry's commitment to a technology may be largely based on short- or mid-term economics, a government commitment to commercialization need not be. The Government may decide that the long-term economics for a technology are a good risk, or that other social values, such as the environment or national defense, dictate development despite current economics. For example, in calling for incentives for the synthetic fuels industry, the Government used the following language to justify the technology on national security grounds:

> The purpose of this title, and the amendments made by this title, are to utilize to the fullest extent the constitutional powers of the Congress to improve the Nation's balance of payments, reduce the threat of economic disruption from oil supply interruptions and increase the Nation's security by reducing its dependence upon imported oil. 18/

If the Government decides to assist private industry in developing new technology, it generally will either assume some of the risk involved or develop the technology itself. This section discusses four areas of risk that influence commercialization: (1) technical risks, (2) economic risks, (3) regulatory considerations, and (4) institutional considerations.

A. Technical Risks.

For technologies beyond the initial research and development stage, there are two primary technical uncertainties: (1) accelerating development, and (2) potential obsolescence. For these technologies, the technical risk involved is not one of basic design, but scale. Development of new technologies under orderly conditions generally requires about 15 years from the time a small-scale pilot plant is started to the time a commercial plant can begin operation. To compress this time frame involves skipping some development steps or reducing testing time. This procedure increases the risks of setbacks down the road.

A second area of technical risk is potential obsolescence. Technological advancement is a dynamic process with continuing improvements being made on existing and even proposed designs. The decision to commercialize a technology generally limits the potential for other designs,

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18/ P.L. 96-294

limiting improvements to incremental ones. The severity of the risks of second, third, and subsequent generations of a given technology depends on the expected life of initial technology, the degree of improvement offered by subsequent generations, and the economics involved.

B. Economic Risks.

Any pioneer venture involves economic risks. These economic barriers generally fall into two catagories: (1) barriers to front-end financing, and (2) uncertainty about long-term economic viability. The severity of these problems depends on the specifics of the technology involved and the structure and expertise of the industry installing it. For example, the structure of the utility industry is fundamentally different than the refining industry; being a regulated industry responsible to a State public utility commission and its own stockholders (if investor-owned), as opposed to the refining industry which must answer primarily to its stockholders and to the actions of its competitors in the marketplace. By the same token, the financing of incremental combustion modifications is fundamentally different than building new combustion devices as a utility could possibly fund low-cost combustion modifications through retained earnings, while construction of a new generating facility would require it to enter the bond market to obtain sufficient financing. Any incentives program would have to recognize these differences and be able to respond in a flexible manner.

A major economic barrier to the commercialization of new technologies is uncertainty regarding project feasibility. This uncertainty includes initial cost uncertainty about a particular plant, and the longer term uncertainty about its overall economics compared with alternatives. Because some of these technologies have yet to be scaled up to commercial sizes, cost estimates are based on estimates and projections, not hard data. When these paper projections are translated into bricks and mortar, the result can make the initial projections appear very conservative.

This question of initial cost uncertainty is related to the second uncertainty regarding long term viability. Just as there is no real life experience in building a commercial size demonstration of some of these technologies, there is no experience in operating one. Questions about reliability, durability, and basic operation and maintainence costs can not be answered until a commercial plant is built. A new design will generally require an extended shakedown period over that of a proven design. Modifications to the design may be necessary. All this can only be projected.

This uncertainty about project feasibility is compounded by another major barrier to commercialization--market uncertainty. The technologies described here provide an environmentally safe way of burning high-sulfur coal as an alternative to low-sulfur coal. However, whether these processes are economic vis-a-vis low-sulfur coal depends partially on the dynamics of the coal market. These dynamics include demand for both high- and low-sulfur coal and the cost of transportation, either by rail, barge, or truck, contractual arrangements, taxes, as well as other concerns such as the situation in the natural gas and oil markets. An acid rain program providing no restrictions on facilities switching to low-sulfur coal would put significant pressure on the low-sulfur coal market and their transportation networks. However, how high the market might go is a manner of speculation. A final market uncertainty would be the potential actions and reactions by companies to technological developments by their various competitors.

C. Regulatory Problems.

The major potential users of the technologies discussed here, utilities, are a very regulated group, primarily by States. Public utility commissions (PUCs) are empowered to set "just and reasonable" rates for electricity consumers. This authority can also entail limits on rates of return, mergers, purchases and sale of properties, issuance of securities, construction of facilities, occurence of competitive bids, company performance standards, allocation of power during shortages, billing rules, and promotion of conservation. Several Federal agencies also have some jurisdiction over utility activities: (1) Federal Energy Regulatory Commission (FERC) which regulates rates for wholesale sales and interstate transmission of electricity; (2) Securities and Exchange Commission (SEC) which administers the Public Utility Holding Company Act of 1935 provision regarding SEC approval of mergers; (3) Nuclear Regulatory Commission (NRC) which licenses the construction and operation of nuclear power plants; (4) Environmental Protection Agency (EPA) which sets pollution standards and monitors compliance; and, (5) Department of Energy which enforces the Fuel Use Act and other DOE regulations.

Perhaps the greatest disincentive under this regulatory network is the concept of "used and useful" which underlies regulation by most State PUCs. Basically, this concept states that electricity consumers should not pay for new facilities unless they are built and generating electricity. Hence, while construction work in progress (CWIP) is proceeding, an allowance for funds used during construction (AFUDC) is set up by the PUC to provide a return to the utility. But this is only an accounting device; real earnings are not received until the unit enters the rate base.

This situation can present a utility with a cash-flow problem if a significant portion of its earnings are AFUDC. Also, bond rating companies tend to downgrade AFUDC earnings as opposed to cash earnings. Hence, in times of low electricity demand, the inability of utilities to place CWIP directly into the rate base can be a significant deterrent to investment in construction, particularly if the technology involved is capital intensive and requires years to construct. Such a deterrent may be entirely appropriate to prevent utilities from constructing plants which consumers have to pay for, but from which they will receive no benefits for several years. The purpose of identifying this problem is not to suggest that CWIP should or should not be placed in the rate base in all cases; it is to suggest that selective inclusion of CWIP in the rate base for pollution control equipment could provide an incentive for utilities to install some of the control technologies being discussed here.

D. Institutional Considerations.

An underlying constraint in introducing significant innovation in industry is resistance to installing new technology which may affect the reliability of its equipment while not producing any revenues. For utilities, the source of this resistance has its roots partially in the 1970s experience with flue-gas desulfurization (FGD) which was essentially forced on utilities by the EPA. Early installations of this technology had significant reliability problems resulting in forced outages and expensive modifications. It also created a significant sludge disposal problem for utilities. Finally, the FGD unit required use of utility funds to build but provided no new generating capacity and, indeed, reduced capacity at plants by a few percentage points. Because many of the technologies discussed here would not have proven records of reliability, utilities' paramount concern about reliability may present a serious barrier to use of these technologies.

Summary

Obviously, the discussion of risk is one of degree. All new development involves risk: accelerated development involves increasing this existing risk even more. Such a situation emphasizes how important the timing of a development process is. In deciding upon accelerated development proposals, a decision has been made whether the urgency of the problem and the potential benefit of the technology in alleviating it is greater than the increased risk involved.

V. INCENTIVES TO ENCOURAGE USE OF NEW TECHNOLOGIES FOR MITIGATING ACID RAIN

If the Congress wished to reduce SO2 and NOx emissions through new technology, the overall objective of an incentives strategy would be to make new technologies more attractive to utilities and other emitters of SO2 and NOx in controlling these emissions than switching their facilities to low-sulfur coal. Developing such a technology-based acid rain mitigation strategy could consist of two steps: (1) develop the technical and institutional base to make these technologies realistic alternatives to fuel-switching; and, (2) provide sufficient economic and regulatory incentives for utilities and other emitters to choose these technologies over alternatives. The following discusses the effectiveness of selected incentives in removing or alleviating the various barriers discussed.

A. STEP 1: Building the Technical and Institutional Base.

The purpose of Step 1 is to bring technical knowledge of and confidence in the new technologies up to the level of current technologies, specifically removing the technical and institutional barriers discussed above. Also, increased technical knowledge would improve the accuracy of cost estimates. This section will discusses three alternatives for achieving this: (1) research, development, and demonstration, (2) information exchanges to improve dissemination of advantages of new technologies, (3) bilateral cooperation with Canada in developing the technologies.

1. Research, Development, and Demonstration. While some of the technologies discussed here are commercially available, uncertainty exists regarding the scaling up some of the other technologies discussed. Hence, some commercial-scale demonstrations would be useful in determining the commercial performance of such technologies under different boiler configurations and operating modes. The expense of such a program depends largely on the technology:

LIMB and Low NOx burners could be retrofitted to plants volunteered by the utility industry while AFBC would require significant new construction.

The object of the program would be to optimize the performance of such technologies under typical operating conditions. Such a program could focus only on the most promising designs with an eye toward standardization where appropriate. Such standardization could be important in reducing cost and increasing reliability of such technology.

However, such Federal involvement in research and development may extend beyond the boundaries of the a sound Federal role as envisioned by the Reagan Administration. The Administration is currently funding small demonstrations (25MW) of LIMB technologies and retrofits of NOx burners, but has continually called for eliminating the AFBC research program.

2. Information Exchanges. Perhaps just as important as technical demonstrations to commercializing new technologies is user confidence in the technologies. Perceived difficulties, valid or not, in using new technologies can prevent users from choosing a technology regardless of its economic advantages. As noted earlier, there is a well-founded reluctance on the part of industry to install some innovations due to its experience with first-generation wet scrubbers. Hence, a purposive and expanded program to include utility representatives in the demonstration of technologies, the dissemination of results, and in informal discussions with various interested parties (environmentalists, government, and other industries) could be useful in breaking down distrust.

3. Bilateral Cooperation. Consideration might also be given to conducting technology demonstrations and information exchanges on a bilateral basis with Canada. Canadian utilities and industry are facing many of the same difficulties as their American counterparts. Such demon-

strations could be beneficial to them as well as to ourselves. Such bilateral cooperation in accelerating development of technology has precedent. During World War II, a cooperative effort was established with Canada for the development of synthetic rubber. A similar procedure could be employed for the effort needed to mitigate acid rain.

Besides technical advantages, cooperation may assist in breaking down some growing distrust between the Canadians and the United States over U.S. commitment to reduce interboundary pollution. This distrust has grown to the point where Canadian representatives have publicly accused the Reagan Administration of stalling on acid rain and have begun discussing acid rain mitigation directly with the Congress. A cooperative research, development, and demonstration program leading to SO2 and NOx reduction might help in reducing such tensions.

B. STEP 2: Commercialization Incentives.

The second step of a proposed acid rain mitigation program could be to encourage use of these new technologies over fuel-switching. For the purposes of this study, we have defined three categories of incentives:

> Direct. Direct incentives are designed to improve the economic feasibility of a technology, reduce project risk, or provide easier access to capital for users. Such incentives may focus on either front-end or long-term operating costs, or both.
>
> Indirect. Indirect incentives are designed to make alternatives to new technologies less attractive, thereby making new technologies more attractive.
>
> Regulatory. Regulatory incentives mandate the use of a specific technology or achievement of a set of goals only attainable with certain technologies.

Each type of incentive works differently depending on the specific technology, the specific incentive, and the structure of the industry.

For the purposes of this paper, we are primarily interested in direct incentives: incentives which will improve inherent economics, reduce project risks, or provide access to capital. This emphasis is designed to maximize the flexibility of the mitigation strategy and minimize long term Government involvement. Some incentives will have a secondary indirect effect of discouraging fuel switching. Of the six incentives discussed, four are primarily financial subsidies, and two involve regulatory changes to improve access to capital.

The primary industry highlighted in this discussion will be the heavily regulated utility industry. As a regulated industry, utilities have some unique problems in regard to financing new technology, particularly access to capital. Hence, while most of the financial subsidies discussed here are equally applicable to regulated utilities as well as other industries emitting SO2 and NOx, the regulatory changes would be only appropriate to the utility industry.

1. Construction Grants. One comprehensive means of commercializing new technologies would be to provide construction grants to assist utilities and other industry in financing these technologies. Such assistance could be partial (sufficient for utilities or industry to choose a specific technology) or total. Such an incentive would have the following beneficial effects:

a. Improve the economic feasibility of the project by funding at least part of the capital costs

b. Reduce project risk by having the government assume some of the financial risk

c. Can provide users with a ready source of capital

d. Could discourage use of fuel-switching by denying grants for that alternative.

Disadvantages to construction grants depend on how they were funded and administered. One possible means of funding a grant program would be from a user fee on consumers of electricity. For example, analysis indicates that the capital costs of a twelve-year eight million ton reduction of SO2 can be funded through either a phased-in 3-mill user fee on fossil-fired electricity generated within the eastern 31 States, or a 1-mill user-fee on all electricity generated in, or imported into, the contiguous 48 States. If new technologies are available to reduce capital costs, a twelve-year, ten million ton reduction can also be funded by these two fee proposals. The average residential rate increase for a 3-mill fee would be under 4 percent and the increase would be under 2 percent for the 1-mill program. 19/ Such percentage increases would generally be larger in the Midwest (which has lower than average electricity rates) and lower in New England (which has higher than average electricity rates). Also, the effective percentage would decline over time as other factors drove up the average cost. The fund would have the effect of levelizing the cost of acid rain mitigation geographically to the benefit of midwestern States. New England States would, in effect, be assisting the Midwest in cleaning up its own pollution; however, the national/international character of the problem could justify such a redistribution. Also, because Midwest residents are more dependent on electricity than New England residents, there would be some extra cost per capita that would be borne by residents of that region.

Besides raising the question of equity, an acid rain "superfund" also raises questions regarding how it would be administered and the duration of the user fee. Answers to these questions would depend on the projected cost of the program, degree of financing the fund was to provide, and the specific technologies funded. While spreading the risk more broadly, such a program would not eliminate the risk involved, and would require the Government to make decisions in an atmosphere of uncertainty, with uncertain results.

19/ Parker, Larry B. Distributing Acid Rain Mitigation Costs: Analysis of a Three-Mill User Fee on Fossil Fuel Electricity Generation. CRS unnumbered report, April 11, 1983; and, Parker, Larry B. Distributing Acid Rain Mitigation Costs: Analysis of a One-Mill User Fee on Electricity Generation in the Contiguous 48 States. CRS unnumbered report, April 22, 1983.

2. Investment Tax Credits. Investment tax credits permit an industry to deduct part of its capital investment from its tax liability, and are employed frequently by the Federal Government to encourage certain investments because of their fairly non-intrusive nature. The advantages of this alternative are:

a. Improves economic feasibility by reducing effective capital costs to industry

b. Reduces project risks by providing benefits in early years

c. Requires no significant active involvement by the Federal Government.

Unlike construction grants, investment tax credits may not be applicable in all cases. First, companies must have a significant tax liability to be offset by the proposed tax credit. This could be a substantial constraint for utilities which may not have the tax liability necessary to take advantage of the proposed credits. Second, public utility commissions may force utilities to provide rate relief from the tax credit rather than allow the utility to retain the savings for capital investment, unless the proposal provides for tax normalization. Finally, it should be noted that investment tax credits will reduce the tax revenues of the federal government at a time when the government is running deficits of record magnitude.

3. Accelerated Depreciation. A standard method of encouraging new investment in fairly risky ventures is to allow accelerated depreciation of the asset. The advantages are:

a. Reduces project risk by providing revenues in the early years

b. Provides shorter payback period resulting in less operating risk

c. Affects the timing, but not the overall take of the Federal Government.

While accelerated depreciation improves the cash flow of a project, it only secondarily improves the project's overall economics or the developer's access to capital by shortening the payback period of the investment. Also, current law already provides some accelerated depreciation benefits to industry. Whether more advantages would provide any additional incentive is unknown.

4. Pollution Control Bonds. Pollution control bonds are a form of loan guarantee to utilities to build pollution control equipment. Their advantages are:

a. Improves economic feasibility by reducing financing costs

b. Provides easier access to capital.

Currently, pollution control bonds are tax-free because they are issued by a State authority on behalf of a local utility. Because they are issued by the State, they take on the character of a municipal bond and hence save the utility about 300 basis points in financing costs, depending on market conditions. However, it is unclear that a State affected by an acid rain mitigation program could extend its "full faith and credit" to cover the potential costs of such a program without possibly increasing taxes.

The credit of the Federal Government could be employed to secure a new pollution control bond specifically designed for an acid rain program. Such bonds could take one of two forms. The bonds could carry a Federal guarantee but no tax advantages thereby saving a utility at least 100 basis points or more depending on the utility's bond rating and market conditions. Alternatively, the bonds could carry both the Federal guarantee and a tax exempt status. Such a bond would save utilities about 300 basis points depending on market conditions, but would cost the Federal Government tax revenues. Both of these proposals could cost the Federal Government money if a utility defaulted, or could strengthen precedent for other types of social spending.

5. Permitting CWIP into the Rate Base for Pollution Control Equipment. Placing construction work in progress (CWIP) into the rate base has been a controversal proposal in some states for many years. The advantages are:

a. May provide easier access to capital

b. Depending upon the circumstances, can have a secondary effect of improving economics by reducing financing costs.

The primary obstacle to allowing CWIP into the rate base has been a fear that it would encourage unwise construction. Yet, if the change were restricted to pollution equipment, this fear might be abated somewhat. However, permitting CWIP into the rate base, while improving the financial condition of utilities and their access to capital, does not directly reduce the risks of commercializing the technologies being discussed here.

6. Consumer-Risk Schemes. Consumer-risk schemes involve ensuring utilities that the cost of new technologies can be passed on to ratepayers and not absorbed by investors. The advantages are:

a. Reduces project risk to investors by transferring it to ratepayers

b. Provides easier access to capital.

As in the case of permitting CWIP into the rate base, consumer-risk schemes are limited to utilities. Also, by transferring cost to ratepayers, there are fewer checks on costs, depending on the specifics of the scheme and the role of the PUC in monitoring it. Hence, the cost of building pollution control equipment might rise under a consumer-risk scheme, perhaps undermining long term public support.

C. Matching Incentives to Technol[illegible] [illegible]lysis of Selected Technologies

As noted previously, an imp[illegible]nt a[illegible] [illegible]ny incentives program is matching the appropriate incent[illegible] to [illegible] [illegible]ology. While the economic aspect of this is very site-specific, [illegible] [illegible]ions can be made regarding which of the above incentives appear [illegible] [illegible]tive for encouraging a specific technology. For the purpose [illegible] this analysis, five technologies, illustrating the range of possibilities, are treated: (1) NOx combustion modifications, (2) physical coal cleaning, (3) LIMB, (4) wet and dry FGD, and (5) AFBC.

Concentrating on the Step II incentives, the impact of each incentive on either the economic feasibility, project risk, or capital availability

TABLE 9: Impact of Incentives on Economic Factors Affecting Commercialization*

	Economic Factor		
INCENTIVE	Economic Feasibility	Project Risk	Access to Capital
1. Construction Grant	1	2	1
2. Accelerated Depreciation	2	1	2
3. Investment Tax Credit	1	2	2
4. Pollution Bonds	2		1
5. CWIP in rate base	2		1
6. Consumer-risk Schemes		1	2

* "1" = primary impact of alternative, "2" = secondary impact of alternative. NOTE: Priorities are relevant only to the individual alternative; no attempt has been made to place priorities across alternatives.

aspect of financing new technology is summarized in Table 9. As indicated, each alternative has specific qualities in regard to encouraging investment. It should be noted that the primary or secondary impact of one incentive on alleviating an economic constraint may be different than the primary or secondary impact of another incentive, in terms of method, degree and effectiveness. Hence, the same general need of different technologies may require different incentives. This is discussed below by technology.

1. Combustion Modifications. The major difficulty with low NOx combustion modifications is convincing utilities of their reliability. In congressional testimony utilities have expressed concern about blowing

up the boiler, corrosion, erosion, and other types of problems. 20/ These concerns must be overcome for widespread commercialization to occur. This can be done by either Step 1 alternatives (particularly demonstrations and information exchanges), or by administrative or regulatory action, if the concerns are determined to be unfounded.

Low NOx combustion modifications represent the most incremental and lowest cost alternative of the technologies being discussed. Cost for retrofitting such equipment on utility boilers range from $0.5 to $5.8 per kilowatt, depending on the specific modification. 21/ For example, to retrofit a 500 megawatt plant with dual register burners providing between 40 and 50 percent removal would cost about $1.25 million. Such an amount could probably be funded through retained earnings, avoiding the need to obtain financing. However, if a utility was required to retrofit several of its plants with such burners, the cost might be sufficient for the utility to require some financing. In this case, a pollution control bond might provide sufficient access to the captial market to install the equipment. Of the other alternatives influencing access to the capital market, the short construction schedule of installing NOx control equipment would tend to negate many of their advantages while construction grants might be considered overkill, given the modest capital costs.

20/ U.S. Senate. Committee on Envirnonment and Public Works. Technical Inquiry on Acid Rain. June 2 and 3, 1982.

21/ Electric Power Research Institute. NOx Regulatory Changes and the Electric Utility. A Guidance Manual for Air Quality Modeling in Planning, Siting, and Emissions Control. September 1981.

2. Physical Coal Cleaning. Coal washing is unique in two respects. First, the technology is proven at commercial scale, and no significant technical questions exist. Second, it is a technology which would be installed primarily by unregulated coal companies, not utilities. For the former reason, a coal washing strategy could be introduced faster than the other altenatives being discussed, indeed, almost immediatuly. In many cases, it is considered a cost-effective alternative when all of the advantages (such as improved boiler performance and lower transportation costs) are included.

For coal washing, a major consideration is the size of the coal company seeking assistance. Most coal washing facilities tend to be large, around 14,000 tons per day input or more. Such a facility is estimated to cost about $30 million and will clean enough coal for a 1,000 megawatt plant. Presumably, smaller facilities could be made for small mine producers who dig nowhere near the 3 million tons per year the above facility could handle, if they have the coal reserves to justify such a long-term investment. Because small mines tend to be somewhat marginal, significant incentives in the form of a construction grant may be necessary to induce purchase.

For larger firms or utilities, the problem would not be as major, the concern being project risk and access to capital. Alternatives such as income tax credits (if the utility has sufficient tax liability) or permitting CWIP into the rate base might be sufficient. In more extreme situations, a pollution control bond might be appropriate.

3. Limestone Injection Multi-stage Burners. The major barrier to LIMB is technical: no commercial size boiler has been retrofitted with the technology. Hence, a program to demonstrate this technology on a

commercial scale would be of paramount importance. It is also a potential research area in which to include Canada, since Ontario and some of the maritime provinces burn fairly high-sulfur coal (some bought from Pennsylvania and West Virginia) and they also need to develop a way of burning it cleaner.

Assuming EPA's cost estimate of $40 per kilowatt installed is accurate, LIMB is projected to be competitive with low-sulfur coal, given current differentials. However, the cost is probably sufficient to require some assistance to utilities for financing and to overcome their reluctance to install a relatively unproven technology. Depending on the financial state of the utility, a pollution bond or even a grant might be necessary to draw utilities from a more proven technology.

4. Flue Gas Desulfurization (Wet and Dry). Unlike some of the other retrofittable technologies discussed here, FGD generally does not claim a cost advantage over switching to low-sulfur coal. However, FGD is a proven technology. Excluding site-specific technical difficulties in some cases, the major constraint on use of FGD is economic feasibility. A construction grant would be the most direct method of improving such economic considerations. However, given the general high cost of FGD compared with the other alternatives being discussed here, administrators of a grant program might choose to encourage FGD only as a last resort.

5. Atmospheric Fluidized Bed Combustion. Unlike the other technologies discussed, AFBC is not currently designed as a retrofittable technology. Instead, significant rebuilding of an existing plant site would be necessary to install AFBC. Combined with its current stage of development, AFBC represents the alternative with the most uncertainty, and therefore, probably the technology in need of the greatest assistance.

Implementation of Step I alternatives would be the first prerequisite to commercializing AFBC. As noted earlier, some significant demonstration work is currently underway. Such effort might consider the possiblities of refurbishing an existing site with AFBC: if the costs can be kept under 50 percent of the initial price, then the AFBC plant would avoid being reclassified as a new source under the Clean Air Act; if not, a waiver could be considered.

For commercializing the technology, a consumer-risk scheme might be weighed along with a partial construction grant depending on the financial strength of the utility. If the technology proves very economic, permiting CWIP into the rate base along with a pollution control bond might be sufficient to encourage use.

D. Developing an Acid Rain Incentives Program: Other Considerations

While research and financial incentives can provide a basis of an implementation strategy, an overall acid rain mitigation program would clearly require a regulatory context in which to function. The overall reduction would have to be mandated, along with a distribution strategy. Some alternatives may require regulatory assistance. For example, coal washing would probably require a regulatory requirement to ensure coal operators of a market for their cleaner, but more expensive coal. The Government may wish to discourage fuel-switching through various regulatory devices.

Hence, a comprehensive acid rain incentives program presents some significant questions regarding the role the Federal Government should take vis-a-vis the various inequalities that would arise under such a program. While major discussion of these questions is beyond the scope of this paper, some of them are identified below.

a. How involved does the Federal Government want to get in order to alleviate the various inequalities (regional and economic) that will occur if economic forces are left to choose acid rain mitigation controls?

b. What relationship would a Federal incentives program have with State PUCs which regulate utilities?

c. What sort of balance does the government want to strike between incentives and regulation?

d. What criteria should the government use in assessing industry's requests for assistance?

e. What would be the proper duration of any incentive program?

Clearly, these are but a representative sample of the questions the Congress and the administrating agency might want to consider in deliberations on a comprehensive incentives program for mitigating acid rain.

VI. CONCLUSION

In avoiding the scrubbing-switching dilemma, the analysis above suggests that much of the problem with new technology may be real and perceived technical difficulties. These technical difficulties result in economic risks and difficulties, particularly in terms of access to capital. (See Table 10)

For many of the technologies discussed in this reader, the real technical difficulties involved are not insurmountable. Indeed, many of them could be resolved and commercial-size demonstrations developed within ten years, if effort is focused on that objective.

TABLE 10: Barriers to Development of Selected Technologies*

	Commercialization Factor			
TECHNOLOGY	Technical/ Institutional**	Economic Feasbility	Project Risk	Access to Capital
NOx Combustion Modifications	1			2
Physical Coal Cleaning			2	1
Limestone Injection Multistage Burners	1			2
Flue-Gas Desulfurization		1		2
Atmospheric Fluidized Bed Combustion	1			2

* "1" = primary problem, "2" secondary problem. NOTE: Priorities are relevant only to the individual technologies; no attempt has been made to place priorities across alternatives.

** Includes both real and preceived technical difficulties.

The economic incentives discussed here are not radical, all of them have been used in various ways in the past. Depending on the overall objectives of the incentives program, they do not necessarily have to involve major government expenditures, but they may. Indeed, if the country decides to levelize the cost of implementing the program through a user fee, then no direct expenditures from the treasury may be necessary. 22/

In the final analysis, the overriding factor in developing an acid rain incentives program is whether a concerted effort with its associated costs by the Government should be undertaken to achieve SO2 and NOx reductions through new technologies. The technologies either exist now or will in the next few years. Most of the technologies are projected to be more than economically competitive with current control technology. The basic decision that has to be made is whether the acid rain problem is serious enough to warrant the effort.

22/ For examples of this alternative, see Parker, Larry B. Distributing Acid Rain Mitigation Costs: Analysis of a Three-Mill User Fee on Fossil Fuel Electricity Generation. CRS unnumbered report. April 11, 1983; and, Parker, Larry B. Distributing Acid Rain Mitigation Costs: Analysis of a One-Mill User Fee on Electricity Generation in the Contiguous 48 States. CRS unnumbered report. April 22, 1983

Governor Rockefeller of West Virginia discusses ways to increase the potential use of coal in the U.S.

"The good news that coal can be burned clean should be proclaimed far and wide by an industry eager to be on the "right" side of an issue that will not go away," says Governor Rockefeller. This article, which appeared in Public Utilities Fortnightly, proclaims that technology is available to reduce pollutants, enabling coal to be burned more cleanly.

Public Utilities
FORTNIGHTLY

Charting a Course for Coal

By The Honorable JOHN D. ROCKEFELLER IV

It was widely assumed, following the onset of wildly spiraling world oil prices in 1973, that the United States' most abundant domestic fuel supply, coal, would soon become the paramount boiler fuel for electric power generation and other industrial uses. While there has been a perceptible shift to coal burning by utilities and industry since then, the process has not been of nearly the proportions expected by the coal industry and others. Here the governor of a major coal-producing state examines the reasons for the coal industry's relative disappointment and suggests what can be done to increase the nation's coal use substantially.

A FRIEND of mine is fond of saying that the future of the coal industry in America is on "automatic pilot." He argues that the fortunes of the industry and the economic prospects of the people of the coalfields are determined by larger forces — the U. S. and world economy, the availability and pricing of other forms of energy, and technological developments. These larger forces, he contends, render industry and political supporters of coal relatively powerless in their efforts to change the course of the industry for the better.

In essence, he would say, the efforts of industry, labor, and political leaders are like the ripples of pebbles tossed in a pond and lost amid the waves of boulders thrown in by outside forces.

There is more than an element of plausibility in his argument. Coal is abundantly available throughout the U. S. The industry is demand-constrained. And in the history of the coal industry, demand has often seemed determined by large outside events and trends seemingly distant and often hard to comprehend.

John D. Rockefeller IV is governor of the state of West Virginia, having been elected to a first term in 1976 and reelected in 1980. He has served as chairman of the President's Commission on Coal and as chairman of the White House Conference on Balanced National Growth. His achievements include his appointment to the National Advisory Council of the Peace Corps and special assistant to the Peace Corps Director. He has also worked in West Virginia with the President's Commission on Juvenile Delinquency and Youth Crime. **Governor Rockefeller** was graduated from Harvard University.

To the people of the coalfields, the very abundance of the domestic energy resource they produce seems to mock the notion that the United States could experience an energy crisis. Yet the boom and bust cycle that has been so much a part of the history of coal is real enough and the factors that produce it must be understood, if the swings in the cycle and their concomitant costs in industry distress and human suffering are to be damped.

The Recent Past

Today's depressed coal market traces back ironically to a tenfold increase in the cost of oil, a major competitive fuel. It would be hard to underestimate the significance of the oil price run-up on the current market for coal. There can be no denying that the tenfold increase in the nominal price of oil from 1973 to 1979 was a major contributing factor in the stagnant economic growth and mounting inflation that characterized our fuel-inefficient society during that time.

The tight money policy of the Federal Reserve Board was adopted in an effort to prevent the oil price shocks of the 1970s from continuing to fuel a broader-based inflation in the U. S. When combined with all-time high budget deficits, this tight money policy brought the highest unemployment and lowest levels of industrial capacity utilization since the Great Depression of the 1930s. Because of the relative size of the U. S. economy and the fact that the economies of the industrialized world are highly integrated, our economic recession becomes theirs as well. The economic downturn in the industrialized nations has cut deeply into the demand for coal. This happened directly insofar as coal is used to produce steam under industrial boilers and indirectly as economic downturn reduced the demand for electricity and steel.

How much of the lost demand for coal will return when the industrialized economies are restored is unclear. Price-induced conservation has permanently reduced energy demand per unit of gross national product. Structural changes away from heavy manufacturing toward higher technology industries are under way, as well. Surely, the severe downturn in economic conditions and changes in the structure of industry in America are examples of the large outside forces my friend had in mind — seemingly beyond the control of the people of the coal industry, yet profound in their impacts.

The notion that coal industry fortunes are in the hands of the fates gains further support from the experience of the electric generating sector. There can be no denying that the slowdown in electricity demand has directly altered the outlook for steam coal demand. Until relatively recently, the accepted rule of thumb among the nation's utility executives was that electricity demand grew at 7 per cent per year with total demand doubling every decade. Many a base-load generating plant was prudently undertaken on the strength of this historical relationship and the belief that it would continue to hold true.

Today, electricity consumption is off substantially. The last year in which growth was near 7 per cent was 1973. Since then, growth per annum has averaged 3 per cent. This past year the growth in electricity demand was less than one per cent.

The excess supply of steam coal on the market today — roughly 100 million tons per annum — is a direct result of coal producers having expanded steam coal production capability to meet the expected demands of a market undergoing fundamental changes. What the electricity industry faces directly and what coal faces indirectly is a downturn in demand from the combined effect of economic recession and price-induced conservation in electrical energy usage.

Nonetheless, coal is expected to continue to increase its share of electricity generation from 52 per cent in 1982 to 55 per cent in 1985 and 60 per cent in 1990. By 1990, electric utilities in the U. S. will be consuming an estimated 845 million tons per year, up 14 per cent from current levels.

"The automatic pilot" — the sense of powerlessness in the face of pervasive change — can be seen in the market for metallurgical coals, as well.

My own state of West Virginia is home for more than half of all U. S. premium-grade metallurgical coal reserves and nearly all low-volatility metallurgical coal reserves. Here, again, technological changes in the manufacture of steel, new practices in materials substitution away from steel, and economic conditions at home and abroad reach back to the metallurgical coalfields.

Lightweight, high-strength plastics are being substituted for steel in traditional steel markets. Technological changes and increased efficiencies in the production of steel have reduced the amount of coking coal used per ton of steel produced. U. S. steel production today is at the lowest level in nearly thirty years. All in all, the market for metallurgical coal is now expected to be flat for the decade ahead.

Taking Responsibility For the Future

Are we then to accept the argument that the coal industry is on "automatic pilot?" Must we in industry, labor, and government who support this industry accept the fatalist argument that we are powerless victims of external forces? I, for one, cannot proceed on that basis. Of course, coal is a basic raw material. Its demand is derived from the demand for energy in general, electricity in particular, and for steel. Recognizing this reality establishes the starting point for our efforts. It should not define the end point of our expectations. It is where we roll up our sleeves, not where we throw up our hands.

People committed to the prosperity of the coal industry can work together to expand markets, keep costs under control, and increase coal's public acceptance. Let me be more specific.

At any given time regardless of external forces, there are two levels of coal demand — the actual and the potential. The potential demand is always greater than the actual demand. The actions that are necessary to move from the actual level of coal demand to the higher, potential level are not in the hands of outside forces, but within our reach. By acting positively to increase coal's potential, we can hope to increase the prospects for the coal industry even in difficult times when external forces are unfavorable to coal.

Both in the U. S. and abroad, the potential for the economic use of coal to substitute for existing or planned oil-fired utility and industrial boilers is substantial. In 1979, the President's Commission on Coal, that I was privileged to chair, identified some 2 million barrels per day of oil that could be economically replaced by coal in the U. S. electricity sector alone. Some of the reconversion candidates that we identified have gone back to coal, but much of that potential demand continues to exist today. At today's oil and coal prices, the stepped-up use of coal in the nation's utility and industrial sectors makes good economic sense. The fuel savings are sufficient to pay back the capital costs of conversion or new construction in a relatively short period of time.

From a national point of view, reconversion would reduce U. S. dependence on imported energy in the years ahead, and U. S. workers would be employed in the domestic energy industry. In theory, at least, despite the slowdown in electricity demand overall, the economic prospects for the utilities, certain energy-intensive industries, and the coal industry all would be brightened by stepping up efforts to replace existing oil-fired steam generation with coal.

In the near term, this would be accomplished through direct combustion of coal. In the years ahead, fluidized bed combustion, coal gasification-combined cycle technology, and coal-based synthetic fuels will help coal play an even larger role. And the case for increased coal use by utilities and industries here in the U. S. is doubly strong among other industrial and developing countries where oil import dependency is far greater.

Obstacles to Be Overcome

If we are to chart a course for the coal industry from the actual demand at any time toward coal's full potential we must recognize and be willing to address the principal obstacles which coal confronts.

Let me be specific about three such areas where the demand for coal can be increased substantially — not through good luck in our treatment at the hands of fate, but through the conscious actions of industry and political leadership.

Utility Regulatory Rate-making Policy

The first has to do with the public utility regulatory process as it involves fuel uses. At today's fuel prices, it is generally accepted to be in the economic interests of utility customers to reconvert coal-capable boilers now burning oil, taking full account of existing clean air laws and regulations. As a result, some utilities have acted to reconvert their coal-capable units. Many have not, despite the fact that reconversion promises to enable utility customers to enjoy more slowly rising costs of generation in the years ahead.

The reason utilities are slow to act on coal reconversion opportunities traces to the regulatory commissions and their treatment of capital costs in the rate base and the workings of the "fuel adjustment clause." Simply put, to be able to pass along more slowly growing utility bills through coal conversion would require hard-pressed utilities to raise capital and to go through the full utility commission rate hearing process. In the absence of a strong consumer movement among ratepayers, it is far less painful from the utility's viewpoint to continue billing customers for high-cost oil, confident of their ability to pass future oil price rises along through the fuel adjustment clause provision.

The way in which our regulatory commissions work to discourage coal reconversion despite the accepted savings from such cost-saving capital expenditures depresses the demand for coal below its potential level. Here the culprit is not to be found in distant global shifts. The causes are closer to home. So, too, are the solutions. We who support coal's future should work to change utility rate commission behavior that discourages coal reconversion where the stepped-up use of coal to replace oil would benefit electricity consumers.

Transportation Costs

Another obstacle preventing coal from achieving its full market potential is rapidly rising rail transportation costs. If the potential markets for coal in utility reconversions and in overseas export are to be realized, the cost advantage coal enjoys over oil must reach the final user and not be siphoned off along the way through monopolistic railroad pricing practices. If coal is to achieve its potential we must keep costs down all along the production path.

Because transportation costs are such an important cost factor, I believe it is time we set to work in the United States to construct competitive coal transportation alternatives to the railroads. In my view, coal slurry lines are the most promising such alternative. Once again, the reduced transportation costs and expanded markets that coal slurry lines promise to the coal industry are not achieved through the whimsy of fate, but by the cooperative workings of people in industry, labor, and government.

Environmental and Health Effects

Finally, there is an aspect of the coal industry in the United States that we who are close to it and who support and depend upon its prosperity must recognize squarely and work to improve. I refer here to the popular perception of the image of coal among potential users and the public in general. In the minds of many the coal industry is synonymous with crippling injuries, labor instability, and environmental degradation. The industry, broadly speaking to include us all, has not put its best foot forward in this area, despite the positive story to be told.

On the subject of safety, for example, the coal commission found, and the National Academy of Sciences subsequently verified in separate investigation, that a substantial amount of underground coal is being mined as safe or safer than surface-mined coal. That is to say, it is being mined with serious accident rates less than half those occurring in the construction industry overall. The fact that industry and labor are working cooperatively to produce coal safely is a story worth telling, not only to the potential users of coal, but to one another in the industry. By aggressively seeking to know why some underground mines can be operated with such exemplary safety records and replicating the causative factors, industry and labor can work together to eliminate one of the negative images that coal now unnecessarily bears.

With respect to the environment, the days of uncontrolled surface mining are gone to history. Today, technology exists to burn coal with insignificant release of sulfur and nitrogen emissions. This is perhaps the best news the industry can offer a skeptical public eager to realize the cost savings and energy independence associated with coal use but increasingly concerned over the long-term effects of emissions from power plants and industrial facilities.

I am convinced that the long-run incremental potential for coal use in the U. S. is closely tied to public acceptance on the environmental issue. New technologies in coal use should be encouraged. The good news that coal can be burned clean should be proclaimed far and wide by an industry eager to be on the "right" side of an issue that will not go away. If coal is frustrated in reaching its full potential because of legitimate environmental resistance, there can be no comfort in bemoaning "external forces." We will simply have ourselves and our own lack of effective public relations to blame.

Summary

I have argued here that the future of coal can be improved by actions we in industry, labor, and government take ourselves — that our prospects are not in the hands of the fates, that coal's future is not on "automatic pilot." There is, perhaps, no better example of this than the efforts of industry and labor leaders to achieve stability in their relationship with one another. There is blame enough to go around with seconds for all involved. But, when the day is done, the potential market for coal in the U. S. and overseas will depend on the growing acceptance, based on experience, that labor and management in the coalfields can resolve their differences within the contract, on an orderly basis, and amid a climate of mutual respect and trust.

This year coal production will reach 837 million tons, an all-time record. The good news is tempered by the expectations that we would do much better. It is easy to charge that, once again, coal has fallen victim to outside forces. Yes, to an extent, but there is much we supporters of the industry can do to move coal when markets are tight. When people say coal's future is on automatic pilot, then it is probably time for coal's friends to reach for the controls.

Major Consulting Firm Takes Stock of Coal

After heavy investment in coal production, coal producers worry about lagging demand, while consumers enjoy low prices. That none-too-rosy industry picture, drawn by Donald R. Gibbons, a coal industry expert at Arthur D. Little, Inc., is in sharp contrast to what was happening in the midseventies when the coal industry expanded in response to the prospect of worsening oil shortages and the Carter administration's drive for energy independence.

Since the 1974 oil embargo, however, demand for coal has only risen at 4 per cent a year — "respectable," says Gibbons, "but well short of the dramatic increases that would have made the producers' investment profitable." Hurt also by strikes and costly governmental regulations, the industry continues to have woes in the 1980s, with the current oil glut curbing demand for coal and keeping down prices.

Companies without long-term contracts to ensure sales of their coal at a reasonable profit, Gibbons notes, have been having difficulty securing an adequate return on their investment in the tough spot market. Some firms have been pushed into bankruptcy, and others have been bought up by well-financed energy companies.

Despite all this, Gibbons is optimistic about the coal industry's future. Worker productivity is rising, and regulations are being eased under the Reagan administration. Over the longer-term, coal's position can only improve since its reserves are huge, while oil supplies are dwindling. As soon as oil prices resume their climb, Gibbons believes coal will look more attractive. Then, the industry's extra capacity will move into production, and the investments of the 1970s will start to pay off.

Another staff member at Arthur D. Little, C. J. Santhanam, points out that an innovative slurry transport technology being developed by the company could deliver cheaper and cleaner coal from America's western deposits. Using liquid carbon dioxide rather than scarcer water to transport coal in pipelines could open up western coal reserves in the United States to utilities that presently use more expensive eastern coal, he says. "Depleted water resources in the western United States are no longer necessary to move coal slurry hundreds of miles to final destinations," according to Santhanam, who has developed a patented method of moving coal through large slurry pipes long distances by using liquid carbon dioxide produced from coal combustion. His slurry system consumes less energy than a coal-water slurry and is environmentally more acceptable, especially in the western states. Using the liquid carbon dioxide for coal transport in pipelines eliminates the need to acquire water rights and the increasing costs associated with the use of water.

At the receiving end of the carbon dioxide slurry line, the liquid carbon dioxide can be completely separated from the coal with minimum energy requirements, Santhanam says, and also that the liquid carbon dioxide can be marketed for enhanced oil recovery from depleted oil fields or even recycled to the mine end of the pipeline for reuse. In June, 1981, W. R. Grace and Company entered into an agreement with Arthur D. Little to develop the technology further and promote its commercialization after completion of pilot tests.

Director of the coal combustion division of the Electric Power Research Institute (EPRI), Kurt Yeager, discusses advancements in technologies that will help to address the problems caused by increased coal burning and subsequent emissions of SO2 and NOx. He focuses on the positive effect these technologies could have on the long range growth of coal use and the coal industry in the U.S.

Along the Technical Front in Coal Utilization

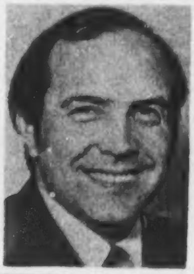
By KURT E. YEAGER

How the electric utility industry and its suppliers respond in this decade to declining demand growth rates, uncertainty over petroleum availability and pricing, stringent environmental controls, lessened capital investment capability, loss of momentum for the nuclear option, decreased governmental funding of technology development, and increased emphasis by domestic coal suppliers on their international market will have important consequences for the industry's power generation capability and structure in the next century. The following article describes technological advances in direct coal utilization which effectively address those challenges.

THE past decade brought new constraints that are reshaping the technical approach of the U. S. utility industry to coal utilization. In the 1970s, increasing electrical rates, restrictive environmental controls, uncertainty over petroleum availability, and skyrocketing fuel costs caused the industry to consider a variety of new coal-based generating options.

Now, in the 1980s, these constraints are compounded by declining demand growth rates, restricted capital investment capability, loss of public confidence in the nuclear initiative, withdrawal of government funding in technology development, and increased emphasis by domestic suppliers on the international market.

Coal now provides over 50 per cent of the electricity generated in the United States. Its importance will continue to grow over the remainder of this century, providing about two-thirds of the nation's energy growth. How the U. S. utility industry and its suppliers respond in this decade to those constraints will have important consequences for the nation's power generation capability and structure into the next century. Certain trends seem clear. For example:



Kurt E. Yeager is director of the coal combustion division of the Electric Power Research Institute and responsible for its principal research and development program on near-term environmental, performance, and reliability issues. Earlier he was director of energy R&D planning for the Environmental Protection Agency's Office of Research. His professional experience has included the development of advanced upper atmospheric sampling systems. Mr. Yeager received an MS degree in physics from the University of California and a BS degree in chemistry from Kenyon College.

Generation capacity will depend increasingly on direct coal combustion. Private sector responsibility for technology improvements will encourage lower-risk, nearer-term options. Priority will be placed on improving the reliability and longevity of existing generating coal capacity to minimize investment.

Technological advances in direct coal utilization which effectively address these challenges are (1) coal quality improvement, (2) improved pulverized coal technology, and (3) fluidized bed combustion (FBC).

Coal Quality

Traditionally, the U. S. utility industry has not given high priority to quality control of its coal feedstock. The general decline in plant reliability and efficiency has, however, led to a reexamination of the entire power generating system from coal supply to ash disposal. The sensitivity of pulverized coal-fired power plant performance and reliability to coal quality has particularly emerged as a major industry concern. As a result utilities are learning that the cheapest coal does not necessarily produce the lowest cost electricity.

A principal facility for assessing the effects of coal quality on power plants is the new Electric Power Research Institute-sponsored coal cleaning test facility

(CCTF) located at the Homer City power station of Pennsylvania Electric Company and New York State Electric and Gas Corporation. Based on commercial scale equipment, the CCTF has a coal handling rate of 20 tons per hour. Unlike other test facilities, it is not dedicated to a single cleaning process but can simulate more than 50 commercial cleaning plant designs. Participating utilities and suppliers deliver 1,000 ton coal samples to the CCTF for cleaning characterization. The results provide a reliable estimate of the commercial cleanability potential of the coal, the effect of cleaning on fouling and slagging, as well as verification of cleaning plant design and cost meeting the company's specific needs.

Coal Quality Assessment

New apparatus to determine coal quality has been developed with the support of the Electric Power Research Institute and is being applied in full-scale utility tests. Called CONAC (for "continuous nuclear analysis of coal"), the method blends the principles of nuclear physics with the practicalities of coal technology. CONAC instrumentation is based on a technique called prompt neutron activation analysis. The technique is already accepted in a wide variety of industrial applications.

CONAC uses a small specimen of radioactive californium (Cf-252) to bombard a coal stream with neutrons. When a neutron is captured by an atomic nucleus, gamma rays are emitted that have a frequency characteristic of the elements involved. CONAC uses radiation detectors to count the number of gamma rays at each characteristic frequency. The number of each is proportional to the abundance of the element associated with that frequency.

CONAC's application range includes determining coal quality at the mine, coal blending, control of coal washing and beneficiation, prediction and avoidance of slagging and fouling, heat management around a boiler, real time heat rate determination, optimal load dispatch, and compliance with flue gas emission limits.

Coal Slurries

Continued oil and gas cost escalation, as well as potential supply interruption, are strong incentives for the conversion of oil-fired power plants to coal. EPRI has conducted studies examining the technical and economic considerations in the several options for achieving this conversion. Based on the results, EPRI is conducting research to develop, demonstrate, and commercialize coal-water slurries (CWS) as a cost-effective, oil replacement fuel for oil-fired utility boilers.

Coal slurries may substantially lower the cost of conversion. The advantage of these slurries over pulverized coal is that they can be transported, stored, and handled with less extensive changes to existing oil-firing facilities.

EPRI, therefore, is focusing its future projects on the rapid development and demonstration of CWS. Various coal feedstocks, coal treatment processes, grinding methods, and stabilizing processes are being studied so that selection and use guidelines can be prepared for the broad range of utility conditions. Combustion tests have been performed in small furnaces and development projects are under way to demonstrate stable and reliable combustion in larger boilers. Utility scale demonstration of CWS combustion is planned by EPRI for 1984-85.

Improved Pulverized Coal Technology

In the next twenty years, the utility industry will be faced with maintaining the integrity of power generation and supply with only a limited number of new plant additions. The difficulty of financing new plants, the cancellation and deferment of many nuclear orders, and the uncertain rate of growth of the industry place increasing emphasis on the availability of existing generation. In particular, many fossil units are approaching the limit of their design lives (typically thirty to forty years) and present a challenge to utilities in extending life without impacting availability.

It is in this current climate that the techniques of incipient failure detection are beginning to flourish. Early warning of component deterioration is seen as an essential part of any predictive maintenance program for fossil plants. Conversely, the absence of such techniques leads to sudden, and often catastrophic, equipment failure causing extended outages for repair and replacement. Even when faced with only one day of additional downtime on a large unit, it is apparent that the capital cost of monitoring equipment is quickly justified.

Diagnostic monitoring is advancing for all areas of the power plant: boilers, turbines, generators, fans, pumps, heat exchangers, but the degree of sophistication of the techniques is not uniform. Some monitoring techniques – vibration signature analysis for example – are well developed and can be implemented into utility maintenance procedures. Others, such as boiler stress and condition analyzers, will require further development and field qualification before across-the-board application to aging fossil plants.

The average station heat rate decreased continuously until the early 1960s. Subsequently, there was little incentive to continue the effort because of the expected increase in nuclear power generation for base-load application and the availability of relatively inexpensive fossil fuels; also additional environmental restrictions required flue gas treatment which had a significant adverse effect on heat rate. A further negative impact resulted from the need to cycle many large fossil plants and run the units at other than base load. This mode of operation typically results in lower component efficiencies and increased heat rate of the units.

EPRI believes that it is important to reverse this trend and has recently carried out a study on the design of fossil plants. As a result, a substantial improvement in heat rate appears possible in new fossil plants. For existing units, EPRI is pursuing a program of on-line performance monitoring and improved instrumentation and testing. The cumulative effect of this effort on the close to 1,000 fossil units currently in service can be a significant reduction in fuel usage as well as immediate economic benefits to the operating utilities.

The trend in unit availability and capacity factor suggests that reserve margins will continue to be of concern in the future. Improved heat rate will be one factor which will assist in improving system reserve margins.

The evaluation of the benefits of higher steam conditions and innovative design concepts must include critical consideration of their impact on unit reliability. Traditional opinion has suggested that supercritical steam-electric plants cannot be expected to produce reliability levels equivalent to those of subcritical plants. Further, the fuel cost savings from high efficiency plants may be more than offset by their increased capital cost and reduced reliability. The EPRI studies concluded that this is unlikely.

U. S. utility experience indicates there is now no statistically significant difference between once through (supercritical) and drum-type (subcritical) unit reliability. Furthermore, we find nothing inherent in the design of supercritical plants which should contribute to lower reliability. The total forced outage hours are about the same for once through and drum units. While, on average, the first large, supercritical power plants achieved poorer availability during the 1960s, the plants built in the 1970s demonstrated results superior to those of drum units. In fact, for plants in the 600- to 825-megawatt-electric range with supercritical pressure and double reheat average availabilities over 80 per cent have been maintained in recent years.

Integrated Environmental Control

Today, environmental regulatory requirements on coal-fired plants involve continuous control of air, water, solid waste, and thermal discharges. This has become a major cost factor in construction and operation, typically 30 to 40 per cent of a coal plant's investment cost. The high-cost, poor reliability, and reduced operational flexibility resulting from the add-on, piecemeal response to these rapidly changing environmental control requirements are changing the historical approach of user, designer, and supplier.

Environmental control in the 1980s has become as much an integral part of the coal-fired power plant system as the boiler or turbine. A design strategy which pursues a more systematic approach to control of all effluent streams may offer capital savings of $100 per kilowatt, increase plant availability by 5 per cent, and improve heat rate by up to 500 kilojoules per kilowatt-hour. These savings occur less from new technology than from assigning single point responsibility for environmental control system design and elevating its engineering priority to a level equivalent with its economic importance.

EPRI's integrated emission control pilot plant (IECPP) at the Arapahoe power station of Public Service Company of Colorado in Denver has the flexibility to test a very wide range of possible equipment configurations. Initially focusing on air, water, and solid waste control technologies, the pilot scale program represents a cost-effective means to obtain critical design and operation information. The overall goal is to produce engineering design guidelines for the selection, configuration, and operation of integrated environmental control equipment under the operating conditions a typical utility might encounter. Emphasis is on the emissions control capability and on the interface among the various components and the power plant to minimize operating, maintenance, and cost requirements.

The IECPP is the first coal-fired pilot system in the United States for investigating different integrated emission control systems and for providing environmental management services to the utility industry. The equipment options provide the unique advantage of sophisticated research flexibility at a size representative of large, commercial installations.

Fluidized Bed Combustion

The electric utility industry also is aggressively evaluating and developing technologies which may ultimately provide less costly and more reliable operation than current pulverized coal systems. These options are intended to carry utility coal utilization beyond the limits inherent in pulverized coal technology. The most important factor in commercial acceptance will be their *demonstrated* capability to be at least as reliable as pulverized coal-fired power plants. Accordingly, priority must be placed on reliability. This requires the construction and operation of large-scale, engineering prototype or pioneer plants over a range of designs and fuels at operating conditions which are representative of the using utility industry.

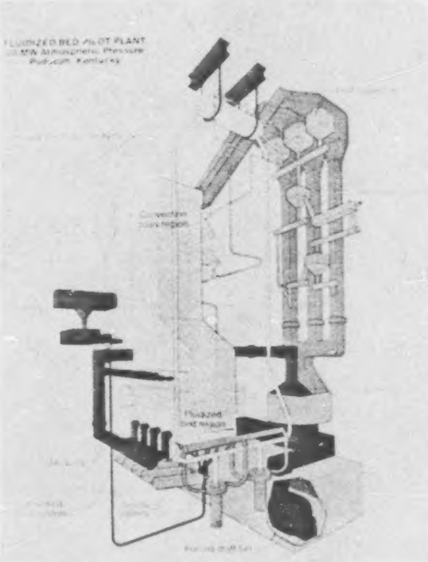
EPRI and the utility industry are accelerating the development and application of fluidized bed combustion as a further *evolutionary* improvement in coal utilization to meet the expanding needs for coal-fired power generation. The improvements which excite this utility interest include reduced sensitivity to fuel quality, thus permitting the use of a much broader fuel supply, from anthracite to municipal refuse, without suffering large losses in efficiency and reliability in a single boiler design. Less cost sensitivity to unit size in a period of load growth and siting restrictions may favor smaller FBC boilers rather than larger pulverized coal furnaces.

A third primary advantage of FBC that may lead to the displacement of pulverized coal boilers is environmental performance. Our experiments with fluidized combustion of coal confirm that it is possible to control sulfur and nitrogen oxides economically without parasitic postcombustion cleanup devices.

Atmospheric Fluidized Bed Combustion

The atmospheric fluidized bed combustion (AFBC) option shows its greatest initial advantage under low-grade or highly variable fuel conditions such as lignite, high-sulfur and high-ash content coal, mine and cleaning plant wastes, and municipal refuse. Since both private and federal projections indicate that a large portion of U. S. coal production growth over the rest of the century and beyond will occur in these low-grade fuels, the AFBC market potential appears substantial. For example, 40 per cent of new utility generating additions during the 1980s will use low-rank coal.

A cost-effective utility scale AFBC design is achievable, and the important hardware questions and alternatives are being resolved. To this end the Tennessee Valley Authority has implemented a 20-megawatt-electric engineering prototype at the Shawnee power station near Paducah, Kentucky. The prototype was built by Babcock and Wilcox. Operation began in May of this year and the EPRI-cosponsored test program will continue through at least 1986. This will provide the basis for 100- to 200-megawatt commercial utility AFBC demonstrations, operational this decade.



The Tennessee Valley Authority's new 20-megawatt atmospheric fluidized bed combustion pilot plant located at Paducah, Kentucky, features a boiler that burns coal and limestone mixed together.

Pressurized Fluid Bed Combustion

The new and dynamic utility climate also influences pressurized fluid bed combustion (PFBC) development goals. The influence arises from the trend toward smaller new unit size and growing utility interest in uprating the capacity of existing units to bring capacity on line at the lowest investment cost. As a result, development emphasis is being placed on turbocharged boilers which can provide shop-fabricated, barge transportable, steam generation modules. These can be rapidly field-erected to provide the desired uprating in unit sizes of 150 to 250 megawatts-electric. This approach will also use coal to replace and increase the capacity of existing oil- or gas-fired plants while minimizing space and environmental control requirements.

The primary physical difference between the turbocharged boiler and the PFBC-combined cycle that has been previously emphasized is the reduction in gas turbine operating temperature. This substantially reduces the development risk and cost, and improves the reliability of the boiler system.

Efforts are now being initiated by EPRI to design engineering prototypes of the turbocharged boiler based on results from operational pilot plants. This approach should permit commercial prototypes to be operational this decade, thus yielding important financial advantages to utilities during the current period of low-load growth and high-construction costs.

Summary

The emphasis in utility coal technology development is on near-term resolution of the reliability, cost, and environmental issues limiting domestic coal utilization. The focus is, therefore, on improvements in conventional pulverized coal power plants and fluidized bed combustion of coal. New apparatus for rapid analysis of coal composition and heat content is being applied to commercial utility installations. Coal quality control has become increasingly important as one element in improving power plant performance and reliability, reducing investment in new capacity, and complying with emission regulations.

Editor's Note: *The foregoing article was adapted from a paper presented by the author at the 12th general meeting of the International Electric Research Exchange held earlier this year in San Francisco.*

ELECTRICAL MANUFACTURERS STUDY HIGH TECHNOLOGY

Staff studies are under way at the National Electrical Manufacturers Association in the area of "high technology" and how it affects the electrical manufacturing industry. NEMA President B. H. Falk has commissioned analyses in six areas: an overview of high technology and its effect on NEMA, the effect of electronics on the electrical industry, technology trends in construction, automated factories, robotics, and fiber optics. Each of the studies will identify market size, major factors in the market, trade association involvement, and other elements that could affect NEMA. These studies are part of NEMA's continuing effort to define high technology in terms of the electrical industry, and clarify the association's role.

— From the "News Bulletin" of the National Electrical Manufacturers Association.

This summary by the Congressional Office of Technology Assessment briefly describes the most promising SOx and NOx control technologies, both available and emerging. Estimation of reduction potential and costs are included when information is available.

C. CONTROL TECHNOLOGIES FOR REDUCING SULFUR DIOXIDE AND NITROGEN OXIDES EMISSIONS

Summary

Acid deposition and ozone are believed to result primarily from the chemical transformation of three pollutants: oxides of sulfur, oxides of nitrogen, and hydrocarbons. This section discusses the most significant control technologies available or under development for reducing emissions of oxides of sulfur and nitrogen. The combustion of fossil fuel is the major source of these pollutants. Three general approaches are used for controlling fossil-fuel emissions: pre-combustion, during combustion, and post-combustion. All three of these approaches have been successfully utilized to reduce emissions from existing sources.

Control approaches also differ widely in the amount of emissions reductions they are capable of achieving and in their cost-effectiveness. Each technology is most cost-effective in a particular range of emission reductions. For instance, the pre-combustion approach of physical coal cleaning is only technically feasible for sulfur dioxide reductions of less than about 40 percent. On the other hand, the post-combustion approach of flue-gas desulfurization is only cost-effective in the 50 to 90 percent sulfur dioxide removal range. The control technique appropriate for a given facility thus depends a great deal on the level of reduction desired. Emission control techniques also vary considerably in their stage of development -- the techniques discussed below are characterized as either "in use," "available," or "emerging."[1]

Controlling Sulfur Dioxide Emissions

I. Precombustion Approaches

Sulfur dioxide is formed when sulfur, an element naturally present in fuel, is oxidized during the combustion process. The greater the concentration of sulfur in the fuel being burned, the greater the production of sulfur dioxide gas. One way to reduce SO_2 emissions is to use fuels with lower concentrations of sulfur. Two important problems with fuel-switching are: 1) low-sulfur coal is more expensive than high-sulfur coal, especially in the east and midwest, and 2) some of the fuel-handling facilities, boiler, and emissions control devices at a power plant have to be modified to burn low-sulfur coal. The potential extent of future fuel-switching is also limited by contraints on the expansion of mining capacity and the availability of transportation facilities to move low-sulfur coals to coal consumers; however, fuel-switching in combination with other emissions control approaches can be a significant and cost-effective means of control.

A second precombustion approach is used to reduce SO_2 emissions by physically or chemically treating coal to remove some of the sulfur it contains. Sulfur in coal exists in two major forms: inorganic and organic. Inorganic or "pyritic" sulfur can be removed inexpensively through physical cleaning. Organic sulfur is chemically bound to the carbon molecules of coal, and can only be removed by a chemical process. Chemical processes are in an earlier stage of development and are more expensive than physical processes, but the sulfur removal potential is higher. Chemical coal cleaning can remove greater than 90 percent of the sulfur in coal, while physical processes can only remove up to 40 percent.

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[1] In-use technologies -- those being sold on a commercial scale in the U.S. that have demonstrated control capabilities.

Available technologies -- those that have been tested and proven but are not currently operational in the U.S. on any significant scale, even though they may be widely utilized in other countries.

Emerging technologies -- those still in the research and development phase, but that have been tested on at least a pilot scale.

II. Combustion Approaches

Currently, no combustion approach is in widespread use for the reduction of SO_2 emissions. Two emerging combustion technologies are the limestone injection multistage burner (LIMB), and fluidized bed combustion (FBC). LIMB is based on the use of staged burner techniques for NO_x control, in combination with dry limestone injection into the combustion chamber for SO_x control. EPA's LIMB research program is expected to continue through 1985, with planning goals to achieve 50 to 70 percent removal of SO_x and NO_x at a capital cost of $30-40/kW. Given continued research, it is possible that LIMB, because it may be retrofitted into existing plants at a competitive cost, may emerge as a particularly attractive control technology option.

Fluidized bed combustion (FBC) involves feeding crushed coal into a bed of inert ash mixed with limestone or dolomite. The bed is fluidized by the injection of air from the bottom of the bed. SO_2, formed during combustion, reacts with the limestone or dolomite to form a solid calcium sulfate, which can then be removed from the boiler. FBC can remove up to 90 percent of the sulfur dioxide; however, costs are very uncertain at this time.

III. Post-Combustion Approaches

Of sulfur dioxide control technologies in use today, flue gas desulfurization achieves the highest reduction percentages for SO_2 removal. Flue-gas desulfurization (FGD) technology removes the sulfur dioxide produced during combustion by spraying the exhaust gases in the stack with a chemical absorbent, typically lime or limestone. This process is popularly referred to as "scrubbing." Of the three types of FGD systems -- wet, dry, and regenerable -- wet processes are most widely used. Wet scrubbers can efficiently remove between 70 to 95 percent of the sulfur dioxide formed during combustion. Dry scrubbing, a much newer technique, has been shown in testing to remove between 60 to 90 percent of the sulfur dioxide formed during combustion. Regenerable FGD systems are currently undergoing major research efforts. The major disadvantage of FGD systems is that they produce large amounts of sludge. A recent government report concluded that in the future the U.S. will produce more sludge from FGD scrubbing than from treating municipal sewage.

Controlling Nitrogen Oxide Emissions

Oxides of nitrogen (NOx) are formed during combustion by two processes: 1) as a result of the oxidation of nitrogen present in the fuel ("fuel NOx") and 2) by the oxidation of nitrogen in the surrounding air ("thermal NOx"). Both processes are controlled by the amount of oxygen present, while thermal NOx formation is also controlled by temperature. The proportion of thermal to fuel NOx produced during combustion varies from fuel to fuel. For coal, EPRI estimates that 20 to 40 percent of NOx emissions are "thermal" and 60 to 80 percent are "fuel."

NOx emissions can be most directly controlled by modifying combustion conditions. While the majority of NOx control techniques focus on the combustion process, post-combustion techniques (flue-gas treatment) are being developed.

I. Combustion Approaches

For technologies in use today, the low-NOx burner achieves the highest percentages of NOx removal. Low-NOx burners, which incorporate features of the staged combustion process, have been shown to significantly reduce the formation of both fuel and thermal NOx at a relatively low cost. During the first stage of combustion, less air is supplied to the burner than is required to completely burn the fuel. Fuel-bound nitrogen is then released, but is converted to N_2 because oxidizing agents are not available. The subsequent addition of air causes the remaining fuel to be burned.

The amount by which NOx emissions can be reduced depends on the type of fuel burned, the type of boiler in use, and whether the plant is new or already exists. Emissions reductions of as much as 50 percent for retrofitting a coal-fired plant, and 60 to 80 percent for retrofitting an oil-fired plant, are achievable using the low-NOx burner technique.

II. Post-Combustion Approaches

Flue-gas treatment (FGT) is an emerging post-combustion process for high levels of NOx removal. FGT has been applied extensively in Japan for use on oil-fired boilers, but has not become as popular as the low-NOx burner in the U.S. because of its high costs for use on coal-fired boilers.

At least 50 different types of flue-gas treatment technologies are available today. Several of them are termed selective catalytic reduction (SCR). In SCR, flue gases are mixed with ammonia (NH_3) and then passed over a catalyst. The catalyst assists in the reaction of ammonia and nitrogen oxides to form nitrogen gas and water vapor. While 90 percent NOx removal is possible using FGT, 80 percent removal is preferable in order to minimize capital and operating costs.

The following Table presents summary information on the major control technologies. Due to site-specific conditions, the removal efficiency levels given should be considered for approximation only.

TABLE 1

OVERVIEW OF CONTROL TECHNOLOGIES

Control Technology	Reduction Efficiencies	Revenue Requirements (Mills/Kwh)	Stage of Development
Sulfur Dioxide			
Fuel Switching	30-90%	0-7	In Use
Physical Coal Cleaning	5-40%	1-5	In Use
Chemical Coal Cleaning	60-85%		Emerging
Wet Flue Gas Desulfurization	70-95%	10-17	In Use
Dry Flue Gas Desulfurization	40-90%	9-15	In Use
Regenerable Flue Gas Desulfurization	70-90%	12-25	Available
Oil Desulfurization			
Indirect	30-40%		In Use
Direct	70-90%		In Use
Limestone Injection Multi-stage Burner	50 90%		Emerging
Fluidized Bed Combustion	<90%		Emerging
Nitrogen Oxides			
Low-NO_x Burner - Commercial	30-50%	0	In Use
Low-NO_x Burner - Developmental	50-80%	0	Emerging
Flue Gas Treatment Without Catalyst	35-40%		Available
Flue Gas Treatment With Catalyst	80-90%		Available

The following article, published in Public Utilities Fortnightly, compares air quality and emission standards for sulfur oxides, nitrogen oxides and particulate matter in the major industrialized countries of Europe, Japan, and the USA. The authors describe what actions have been taken in these various countries to curb pollutants and suggest that the Japanese and California experiences may be a good guide for possible future pollution control requirements.

Environmental Protection from Power Generation: An International Overview

By CARLOS MURAWCZYK and KEN M. MOY

This article concerns itself with a comparative description of developments which have taken place or are taking place in Western Europe and Japan, as well as in the United States, by way of efforts to mitigate the air-quality effects of increased energy production by fossil-fueled power generating plants.

IN 1970s certain industrialized countries of the world experienced a surge in construction of power generation plants together with a proliferation of laws and regulations to control the pollutants emitted by them. This article will seek to bring into focus the various national concepts on pollution control adopted by countries in Western Europe, Japan, and the United States.

Throughout the industrialized world, and more specifically among those countries making up the Organization for Economic Cooperation and Development,[1] significant efforts have been made to maintain and improve the environmental quality of life; especially in areas related to protection of public health and welfare, while at the same time policies were drawn to alleviate the energy crisis. These policies differed from France's commitment to energy independence via nuclear technology, to maximization of energy conservation in the Scandinavian nations, to a mandate in the U. S. to shift back to coal for steam and power generation.

The following evaluation concerns itself with those developments which have been or are taking place in Western Europe and Japan vis-a-vis the United States in mitigating one of the effects of increased energy production by fossil-fired power plants—air pollution.

Carlos Murawczyk manages the policy analysis group in the Advanced Engineering Department of United Engineers and Constructors Inc. He specializes in assessing regulatory and economic impacts involved in licensing steam-electric generating facilities and industrial complexes. **Mr. Murawczyk** has a BS degree in zoology and chemistry and an MS degree in environmental sciences.

Sulfur Dioxide Control

The major point sources of man-made atmospheric contamination are industrial and utility complexes that utilize large quantities of fossil fuels (coal or oil) to generate process steam or electricity. In Western Europe the reduction of the pollutants in coal has been achieved through coal washing, whereas this method of SO_2 control has not been as commonly utilized in the U. S. Coal washing has been effective in Europe since the average European coal has a low-sulfur content (1.35 per cent) in its natural state and has a large fraction of the sulfur bound in the pyritic form which is easily removed by washing. In the United Kingdom, approximately 50 per cent of the steam coal is washed at the present time, and the Federal Republic of Germany has announced its intention to wash all coals to achieve an average sulfur content of one per cent or less by 1985.[2]

It is interesting to note that the original goal of coal washing in Europe was to reduce the ash content of the coal, and thereby obtain a more uniformly combustible fuel. Sulfur reduction was incidental to the washing process.

Certain countries in Europe — e.g., United Kingdom, Federal Republic of Germany, and France — in addition to limiting the sulfur content of the fuels used, specify that the stack serving a facility must be of sufficient height so that ground-level concentrations will meet ambient air-quality standards.[3] This approach is quite different from that in the U. S. where Congress has determined that no credit for reduced ground-level concentrations will be allowed once a stack height exceeds that necessary for "good engineering practice"; i.e., that height necessary to avoid aerodynamic downwash effects. It should be noted that reliance on tall stacks and low-sulfur fuels in Western Europe will not result in sufficient mitigation of air pollutants in the future. Some of the environmental problems which must be faced in relying on tall stacks as an SO_2 mitigative measure are:

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[1] OECD member countries are: Australia, Belgium, Canada, Denmark, Finland, Italy, Japan, Luxemburg, the Netherlands, New Zealand, Norway, Portugal, Spain, Sweden, Switzerland, Turkey, United Kingdom, and the United States.

[2] "The Long Range Transport of Air Pollution — A Review of European Experience," by R. A. Barnes, *Journal of Air Pollution Control Association*, Vol. 29, No. 12, December, 1979.

[3] "The World's Air Quality Management Standards," by W. Martin and A. C. Stern, U. S. Environmental Protection Agency, Office of Research and Development, Washington D. C., October, 1974. EPA-650/9-75-001/a; "Air Pollution Constraints on Increased Coal Use by Industry," by E. S. Rubin, *Journal of Air Pollution Control Association*, Vol. 31, No. 4, April, 1981.

1) The increasingly adverse effects of acid rain, especially in the Scandinavian countries.
2) The anticipated increase in SO_2 emissions throughout Western Europe as a result of increased energy consumption.
3) The drawing down of the supply of low-sulfur coal which can be washed.
4) The increasing cost of fuel desulfurization relative to that for flue gas desulfurization.

Flue gas desulfurization facilities (FGD) are in limited use in Western Europe. In 1978 the Federal Republic of Germany had three FGD systems in operation (lime-sludge, lime-gypsum, carbon-acid or sulfur), France one (ammonia-sulfur), Sweden one (lime-sludge), and Norway one (seawater). The United Kingdom, although the early (1933) developer of the lime-limestone SO_2 removal process for a coal-fired power plant, has only one FGD system in operation. Japan had 58 FGD systems installed as of 1979, and the United States has over 90 systems presently in operations.[4]

In order to bring the various air-quality standards in the member countries of the European Economic Communities (EEC) into harmony, the council of the EEC has adopted international ambient air-quality standards. The council has directed the member states to take appropriate measures to comply with these standards, by April 1, 1983.[5]

Table 1 shows the EEC air-quality standards compared to those in a few member states, and in Japan and the U. S. The EEC standards for SO_2 and particulate matter are, in general, more stringent than those presently in effect in most member states and the U. S. It should be noted that EEC standards take into account the synergistic effects of combined SO_2 and particulate matter concentrations. It appears, therefore, that the motivational force and basis have been established for a much greater utilization of FGD technology in Western Europe.

In contrast to the European air pollution regulations, which are air quality directed through whatever means practical, those in the U. S. are technology forcing; that is, utilities are required to use the best available control technology which has been demonstrated to be commercially available to achieve regulatory requirements on new plants.[6]

The country with the largest diversity of FGD processes and perhaps the most advanced technology is Japan. There were 58 major FGD systems operating in Japan as of 1979. Most of these systems were installed in oil-fired industrial boilers and are regenerable (recycle the scrubber reagent) or by-product recovery (produce a marketable product such as gypsum) systems. A number of FGD systems are installed on coal-fired power plants in Japan. Several of these plants burn low-sulfur coal

TABLE 1
SO_2 AND PARTICULATE AIR-QUALITY STANDARDS
IN SELECTED INDUSTRIAL COUNTRIES[a]
(Micrograms Per Cubic Meter)

	Sulfur Dioxide (SO_2)			*Particular Matter (PM)*		
Country	*Annual*	*Daily*	*Other*	*Annual*	*Daily*	*Other*
EEC	80 if PM>40; 120 if PM<40	350 if PM<150; 250 if PM>150	130 if PM>60;[c] 180 if PM<60[c]	80	250	130[i]
France		250			350[h]	
Germany			140[f] 400[d]			200[f,m] 400[f,m]
Italy		260	650[d,h]		300	750[h,k]
Norway		200	60[c] 400[j]		120	40[c]
Sweden		300[e] 200[f]	60[c] 750[i]		120	40[c]
USA	80	365	1,300[b,l,m]	75 60	260 150[b,m]	
Japan		100	260[i]		100	200[i]

FOOTNOTES:

[a]As of 1978 and 1980.	[h]Special zones.
[b]Secondary standard.	[i]One-hour average.
[c]Six-month average.	[j]Winter average.
[d]One-half-hour average.	[k]Two-hour average.
[e]Maximum value.	[l]Three-hour average.
[f]Long-term average.	[m]Various state standards may also apply.
[g]Short-term average.	[n]Also, half this value for PM>10 microns.

REFERENCES:

1. "Limitation des Emissions de Pollutants et Qualite de l'Air, Valeurs Reglementaires dans les Principaux Pays Industrialises, Specifications en Vigueur en 1978," Institut Francais de l'Energie, Paris, July, 1978.
2. Council of European Communities Directive on Air Quality Limit Values and Guide Values for Sulphur Dioxide and Suspended Particulates, Brussels, July 15, 1980.

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[4] See Footnote 2.

[5] Council of European Communities Directive on Air Quality Limit Values and Guide Values for Sulphur Dioxide and Suspended Particulates, European Economic Community Council, July 15, 1980 (80/779/EEC-OJL229, August 30, 1980).

[6] Clean Air Act—Prevention of Significant Deterioration (PSD) Regulation, Title 40, Code of Federal Regulations, Part 52.21.

(0.6 per cent sulfur) and in addition remove 85 to 95 per cent of the SO_2 from the flue gas. Regenerable and by-product recoverable systems are encouraged in Japan since land for disposal of sludge and ash are at a premium.

TABLE 2

SO₂ EMISSION STANDARDS FOR SOLID FUEL COMBUSTION IN SELECTED INDUSTRIAL COUNTRIES

Country	*Emission Standard*
EEC	None
United Kingdom	Specified stack height to limit short-term ambient concentration (nominal three-minute average SO_2 level 0.17 ppm).
France	Stack height criteria and lowest per cent sulfur fuel if ambient SO_2 level 1,000 micrograms per cubic meter. In Paris and North zone, 2.2 lb SO_2 per million Btu. In Rhone zone, 1% sulfur in fuel.
Germany	For 1,100 Mw thermal input, 1% sulfur in fuel. For 1.100 Mw thermal input, 1.7 lb SO_2 per million.
Norway	None
Sweden	None
USA	For 73.3 Mw thermal input, see state standards. For 73.3 Mw thermal input, 1.2 lb SO_2 per million Btu and 90% SO_2 removal; if SO_2 emissions 0.6 lb per million Btu, only 70% SO_2 removal required.
Japan	Allowable SO_2 emissions determined from both emission rate equation: SO_2 (m^3per hour) = $K.10^{-3}He^2$, K = 1.17 to 17.5 depending on district, He — Effective stack height, and maximum permissible ground-level SO_2 concentration: 0.002 ppm to 0.030 ppm depending on district.

REFERENCES:

1. "Limitation des Emissions de Pollutants et Qualite de l'Air, Valeurs Reglementaires dans les Principaux Pays Industrialises, Specifications en Vigueur en 1978," Institut Francais de l'Energie, Paris, July, 1978.
2. "The Clean Air (Emissions of Grit and Dust from Furnaces) Regulations 1971," No. 162, United Kingdom Department of Environment, London.
3. U. S. Environmental Protection Agency Regulations on Standards of Performance for New Stationary Sources, Title 40, Code of Federal Regulations, Part 60, Subpart Da, June 11, 1979.

Japanese SO_2 emissions are controlled by regulations which are structually similar to those in Western Europe. An examination of the SO_2 emission standards for Japan shown in Table 2 reveals that, although allowable SO_2 emissions are proportional to the effective stack height, they are also constrained by the maximum ground-level SO_2 concentration permitted in the plant's district. The permissible ground-level SO_2 concentration can be as small as 0.002 parts per million (five micrograms per cubic meter).

Of the 90-plus FGD systems in the United States only seven are of the regenerable type. The U. S. has recently expressed a great deal of concern over industrial hazardous and toxic waste disposal practices. The Environmental Protection Agency has announced its intention to review the methods electric utilities use for ash and FGD sludge disposal, even though these are presently exempt from consideration as toxic wastes.[7] It is foreseeable that in time regulations encouraging or imposing utilization of regenerable or by-product recovery FGD systems in the U. S. may be enacted. Regenerable magnesium oxide FGD systems and the Wellman-Lord FGD systems have already been installed by some U. S. eastern and western utilities. The capital costs of these regenerable types of FGD systems are approximately $280 per kilowatt and $340 per kilowatt, respectively, while the cost of the most used limestone-based throwaway FGD system is $190 per kilowatt in 1982 dollars. Operating costs for the magnesium oxide, Wellman-Lord, and limestone FGD systems are approximately 5.5 mills, 8.1 mills, and 7.7 mills per kilowatt-hour, respectively, in 1982 dollars. Annual revenue costs are, respectively, 12.9, 17.1, and 12.7 mills per kilowatt-hour.[8]

TABLE 3

PARTICULATE EMISSION STANDARDS FOR SOLID FUEL COMBUSTION IN SELECTED INDUSTRIAL COUNTRIES

Country	*Emission Standard*
EEC	None
United Kingdom	For new power plants > 150 Mw thermal input, 115 mg/Nm₃, and Ringelmann Number ≤ 2.
France	For plants ≥ 108 Mw thermal input, 0.08 lb per million Btu and Bacharach Number ≤ 6 (except ≤ 4 in special zones).
Germany	For water tube boilers (≥ 500,000 m^3 per hr), 100 mg/Nm^3 (lignite) and 150 mg/Nm^3 (bituminous coal), and Ringelmann Number 1.
USA	For new steam-electric generating units, 0.03 lb per million Btu and Opacity ≤ 20%.
Japan	For new boilers, 400 mg/Nm^3 if burn lower-grade coal, 200 mg/Nm^3 if burn other grades of coal.

REFERENCES:

1. "Limitation des Emissions de Pollutants et Qualite de l'Air, Valeurs Reglementaires dans les Principaux Pays Industrialises, Specifications en Vigueur en 1978," by P. Jarrault, Institut Francais de l'Energie, Paris, July, 1978.
2. "The Clean Air (Emissions of Grit and Dust from Furnaces) Regulations 1971," No. 162, United Kingdom Department of Environment, London.
3. U. S. Environmental Protection Agency Regulations on Standards of Performance for New Stationary Sources, Title 40, Code of Federal Regulations, Part 60, Subpart Da, June 11, 1979.

Particulate Control

A comparison of the regulatory requirements for the control of fly ash emissions in the U. S., Western Europe, and Japan reveals that U. S. coal-fired plants are the most stringently controlled. Table 3 shows that the European and Japanese standards for particulates are based on a particle size of ten microns or less for human health reasons. In the U. S., the Environmental Protection Agency has also embarked on the development of an ambient air-quality standard for fine particulate matter (less than ten micron diameter). The purpose of this standard is to protect the public more adequately from particles which can reach the oxygen exchanging tissue in the lung. At this time, the effect of such a standard on the allowable particulate emissions from a coal-fired boiler remains to be seen.

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[7] "Effluent Limitations Guidelines, Pretreatment Standards, and New Source Performance Standards under Clean Water Act; Steam Electric Power Generating Point Source Category," *Federal Register*, Vol. 45, No. 200, October 14, 1980.

[8] United Engineers and Constructors cost estimates.

The most widely used means of particulate emissions control in the U. S. has been the electrostatic precipitator, although fabric filter baghouse collectors are increasingly being specified for new coal-fired plants. This change arose almost concurrently with EPA's revision of the federal standards of performance for new stationary sources regulations specifying 99-plus per cent collection efficiency for particulates from fossil fuel-fired utility boilers. This level of efficiency is generally more easily attainable by baghouse collectors and at less cost for low-sulfur (high-resistivity) western coal. The operating costs, including bag replacement costs, of a baghouse collector have also been shown to be comparable to those of a high-efficiency electrostatic precipitator at several electric utility coal-fired boilers. The capital and operating costs of a precipitator or baghouse collector capable of meeting the U. S. particulate emission standard of 0.03 pounds per million Btu are $45 to $90 per kilowatt and 1.5 to 3.2 mills per kilowatt-hour, respectively, in 1980 dollars.[9]

Control of Oxides of Nitrogen

In addition to emissions of SO_2 and particulate matter, fossil fuel-fired power plants are major emitters of oxides of nitrogen (approximately 95 per cent nitric oxide (NO) and 5 per cent nitrogen dioxide (NO_2)). Only Japan and the U. S. have emission control requirements on oxides of nitrogen (NO_x). (See Table 4.) In the U. S. boiler manufacturers have used combustion process modifications in order to guarantee that new, coal-fired steam generation will meet federal New Source Performance Standards (NSPS) limits for NO_x. These modifications include one or more of the following:

1) Staged Combustion
2) Low Excess Air
3) Reduced Heat Release Rate

NO_x emission controls more stringent than those presently regulated by NSPS regulations may be required under certain circumstances. California Air Pollution Control Districts require that NO_x emissions from a new source be offset in some manner where the ambient air-quality standards for photochemical oxidants are violated. Combustion process modifications, as outlined previously are inadequate to meet NO_1 emissions limitations more stringent than those in the current NSPS regulations. Postcombustion control techniques appear to offer a feasible alternative.

Of the several postcombustion techniques that have been investigated, selective catalytic reduction (SCR) of NO_2 by ammonia (NH_3) has shown the most promise.

Selective catalytic reduction of NO_x by NH_3 is based on the preference for the reaction of NH_3 with NO_x rather than the other flue gas constituents. The SCR process requires only a catalyst in a reactor, and an ammonia storage and injection system. The reactor operates in a temperature range between 570 and 850 degrees Fahrenheit. At the present time, the SCR process has been demonstrated, on a commercial scale only on oil- and gas-fired boilers. This process is being applied to coal-fired boilers presently on the pilot plant scale in both the U. S. and Japan. For a coal-fired utility boiler, factors such as poisoning of the catalyst by the uncleaned flue gases, catalyst life, and reactor effectiveness still need to be investigated.

Finally, the cost of the SCR process on a new coal-fired utility boiler merits discussion. For two 500-megawatt boilers ordered in 1985, the capital and operating cost (in constant 1980 dollars) for the SCR process are estimated to be $45 to $85 per kilowatt and 6.2 to 8 mills per kilowatt-hour, respectively.[10]

TABLE 4

NITROGEN OXIDES EMISSION STANDARDS FOR COAL COMBUSTION IN SELECTED INDUSTRIAL COUNTRIES

Country	*Emission Standard*	
USA	Coal Classification	NO_x (as NO_2) Limitation
	Subbituminous	0.5 lb per million Btu
	Bituminous	0.6 lb per million Btu
	Dakota or Montana lignite	0.8 lb per million Btu
Japan	480 ppm at 6% O_2 (~ 0.8 lb per million Btu)	
EEC	None	
United Kingdom	None	
France	None	
Germany	NO_x emissions to be reduced as much as possible by technical means	

Conclusions and Recommendations

Present-day air pollution control technology for fossil fuel-fired power plants varies from the use of tall stacks and low-sulfur fuels in Western Europe to the development of expensive flue gas cleaning equipment in the U. S. and Japan. The various approaches to attaining and maintaining specified air-quality levels adequate to protect the public and the environment is dictated by economic as well as legislated approaches. This is clearly evident in the sizeable number of FGD systems which have been installed within the past five or six years in the U. S. and Japan but not in Western Europe.

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[9] "Control Technology: Status of EPRI Research," by G. T. Preston, Edison Electric Institute—Envirosphere Company fourth annual conference, Washington, D. C., October 26-28, 1981.

[10] See Footnote 9.

Further development and application of flue gas cleaning systems can be expected in progressive industrialized countries of the world, especially in view of recent progress in gas cleaning equipment performance and regulatory history. U. S. regulations require a review of emission standards for fossil fuel-fired plants every ten years. Should the economy rebound and environmental concerns heighten, fossil-fired power plants will again be viewed by an aroused public as major contributors to acid rain, visibility impairment, and heavy metals in the atmosphere. The Japanese and California experience may portend a requirement for further pollution controls in U. S. plants.

Program of Fellowships in Renewable Energy Engineering Created

Responding to a growing crisis in engineering education and a shortage of qualified engineers with experience in renewable energy, a "fellows in renewable energy engineering program" is being established as a partnership of the College of Engineering at the University of Hawaii, the university-affiliated Hawaii Natural Energy Institute, and the Renewable Energy Institute in Washington, D. C. The fellows program involves corporate and foundation endowment of teaching and research positions to advance the development of renewable energy technologies.

This project was developed through the joint efforts of U.S. Senator Spark Matsunaga, a longtime supporter of renewable energy development and the staffs of the institute and the University of Hawaii. Initiated as a seed project, the fellows program matches a full professor's salary and rank (provided by the University of Hawaii) with additional funding increments provided by the program's sponsors to attract the finest in engineering faculty to work on critical renewable energy engineering problems in ideal natural laboratories.

With superb conditions for wind power, direct solar energy, geothermal energy, ocean energy, and biomass production, Hawaii is thought to be an excellent location for this program. Throughout the islands, virtually every renewable energy technology is in some stage of advanced development or commercialization. Hawaii currently imports more than 90 per cent of the energy it consumes, mostly from member nations of the Organization of Petroleum Exporting Countries, so there is a strong desire to move away from oil dependency to renewable energy. Renewable energy sources are the only locally available options because there are no fossil fuel deposits located anywhere in the state.

The fellows program is expected to permit growth of a constellation of renewable energy engineering faculty positions, matching private sector contributions with university funds set aside for this purpose. The program will be guided by a distinguished board of advisors, chaired by Senator Matsunaga, and will be governed by a steering committee representing the financial and institutional sponsors.

Joining as the initial corporate sponsors are the General Electric Company, Chevron Corporation, and Atlantic Richfield Company.

Printed in the Electric Power Research Institute (EPRI) Journal, this article describes the technological refinement and determination of the Japanese to use pollution control systems to control industrial and utility emissions. By comparing developments in the U.S. and Japanese auto industry, the article voices concern that similar domination by the Japanese of pollution control technology will occur. U.S. unwillingness to refine and use technology originally developed here has cost us preeminence in the field of pollution control industries. technologies which could have been exported to other industrialized countries and emerging nations.

Japanese Challenge in Coal Combustion Technology

About 10 years ago, the Japanese utility industry used no coal. Now Japan rapidly becoming a world leader in applying the state of the art in pulverized-coal-fired electricity generating technology.

"The rapid development of coal combustion technology and facilities in Japan is both exciting and disturbing," reports Kurt Yeager, director of EPRI's Coal Combustion Systems Division. "Exciting in that utility technology developed originally in the United States is being improved and implemented successfully. Disturbing in that the Japanese have seized the initiative and are able to make technical and commercial progress where the United States is unwilling or unable to do so."

The U.S. market for coal-fired systems fell off dramatically in the 1970s and is expected to recover only partially in the next decade. As U.S. suppliers turn toward the growing international markets, particularly those in the Third World, they will find their technical leadership challenged by Japan.

Yeager's observations stem from a visit to Japan with other senior EPRI management as part of scheduled biennial meetings with managers at the Central Research Institute of the Electric Power Industry (CRIEPI). In addition to discussions at CRIEPI, Yeager's visit included stops at facilities of Japan's Electric Power Development Co., Tokyo Electric Power Co., and Mitsubishi Heavy Industries.

"I observed no technological or management breakthroughs during my visit," says Yeager. "The Japanese are simply applying the best technology, much of it pioneered in the United States, with more concerted attention to design, engineering, operating, and maintenance practices than is typical in this country. At this point, I judge they have the capability and, just as important, the commitment to design, build, and operate coal-fired plants as well as anyone in the world. This is evidenced by their competitiveness in the international market and by their domestic accomplishments."

The approach taken by the Japanese has been to carefully examine conventional coal-fired power plant experience throughout the world and to objectively integrate the best elements with no constraints or biases in terms of traditional practices and procedures. In some areas, such as instrumentation, NO_x control, and coal pulverizers, they have made technology advances of their own. However, their basic advantage results from better integration and attention to detail in implementing technology.

In this, they seem generally content for the time being to leave development of revolutionary technology to others. Advanced coal-based options are of interest in Japan, but primarily for longer-term application—probably beyond this century. In the interim, their analyses indicate that under many circumstances, conventional coal-fired generating technology can produce electricity at costs and with efficiencies equivalent to those projected for the more advanced options.

The Japanese emphasis could have important repercussions for the preeminent position of U.S. power plant manufacturers and suppliers. In the same manner as Japanese automakers moved into the U.S. automobile market, their strong focus on integrating and improving the best of conventional technology could enable them to begin chipping away at the market for power generation and auxiliary equipment in this country, as well as internationally through licensing or supply. This rapidly provides the experience and market base for leadership in the more innovative advances in technology.

All plants to be built in Japan will function under supercritical steam conditions (i.e., at pressures above 3450 psi; 23.8 MPa) and incorporate a high degree of fuel flexibility, improved instrumentation, integrated environmental controls, and various other technology improvements at both component and system levels. Because of rate structures and higher fuel costs in Japan, emphasis is also being placed on reducing lifetime costs as opposed to minimizing capital investment. Notwithstanding this, capital costs are competitive with those of conventional pulverized-coal-fired plants now being installed in the United States.

According to Yeager, the key to Japan's success in advancing conventional coal-fired technology is its willingness to objectively evaluate equipment options and design, operating, and maintenance practices on the basis of value, not past practices and tradition. The Japanese are willing to regularly and critically evaluate their practices in an effort to pinpoint potential improvement.

On returning from his second visit to Japan, Dan Giovanni, program manager in the Coal Combustion Systems Division, pointed out that there has never been any serious concern in Japan, as there has been in the United States, that reliability and efficiency might be in conflict. Once the Japanese utilities recognized the efficiency advantages that could be realized with supercritical steam conditions, they aggressively pursued them. After having identified the best available technology, they purchased it on the assumption that when it was installed and in place it would represent the best available power plant.

A second important factor in their success is that Japanese manufacturers have taken an aggressive position in product development and do so in close cooperation with their utility clients. In return, Japanese utilities invest heavily in providing development and demonstration facilities and appear to have more effective control over development of technology for their use. Government support in this process is also critical, with government, industry, and utilities working together rather than at odds with one another.

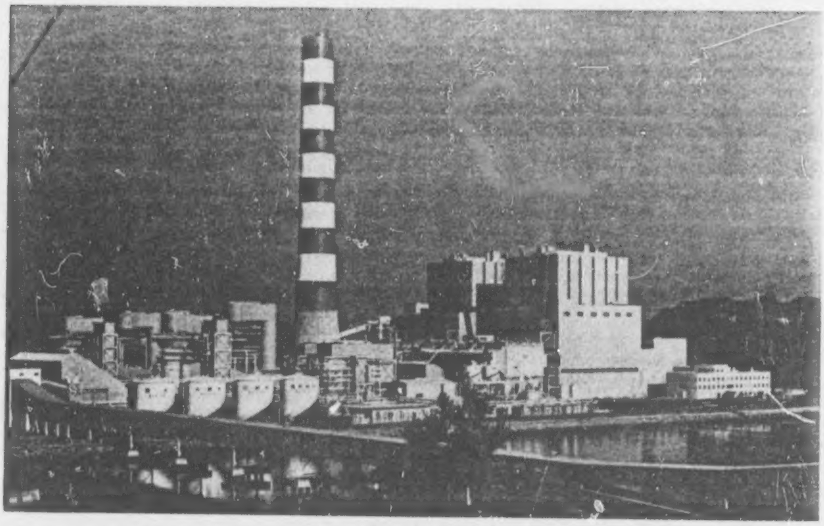
"Their 'secret,'" contends Yeager, "seems to be a higher sense of responsibility to do the job right than we in the United States generally now apply. They are convinced that their commercial success and resulting national survival depend on this dedication. Also, key decisions are made in a more timely manner, based on consensus. This approach is dramatically reflected in the scheduling of siting and licensing procedures." The results attainable by this philosophy of accord are perhaps best reflected in the siting of the Electric Power Development Co.'s Matsushima Power Station, which was built on a pristine and culturally significant island off the coast of Kyushu near Nagasaki. Commissioned in 1981 five years after license application at a cost only slightly above that of conventional coal-fired plants in the United States, Matsushima is widely believed to be one of the most advanced coal-fired power stations in the world.

It consists of two 500-MW (e), 3600-psi (24.8-MPa), single-reheat, supercritical units of licensed U.S. design; combustion NO_x control; hot-side electrostatic precipitators; and an SO_2 scrubber capable of handling the full flue gas flow for each generating unit.

Matsushima is a baseload plant. Net heat rate is about 9200 Btu/kWh, and the load change rate is nominally 1% a minute, although the plant has been tested up to 3% a minute. It is designed to use coal from Australia, China, Canada, and the United States, and it has totally enclosed coal-handling and -unloading facilities for ships of up to 60,000 tons capacity.

In addition to operating at supercritical steam conditions and having the flexibility to use a variety of different coals, Matsushima has a number of other key characteristics.

□ Improved instrumentation and human factors design. In this regard, there are two independent control rooms that present more data in a far more readable form than do their U.S. counterparts. Also, plant control appears to be more precise, and diagnostic instrumentation for maintenance planning is much more extensive.



Japan's current target for developing conventional pulverized-coal-fired technology is a 5000-psi (34.5-MPa) supercritical plant operating at 1200°F (649°C) with two reheats to 1100°F (593°C). The plant would also incorporate reduced power consumption by the auxiliary equipment and a higher performance by the air preheater. This is projected to provide a 7% improvement in efficiency at the busbar (from 36.7% to 39.3%) and to result in a coal combustion reduction of 160,000 tons a year for a 1000-MW (e) plant.

The capital cost increase over a 1000-MW (e), 3500-psi (24-MPa), 1000°F (538°C) single-reheat plant is estimated at $45 million ($45/kW), with annual fuel savings of about $12 million a year ($3/MBtu). For lower-risk steam conditions of 4500 psi (31 MPa), 1050°F (560°C), the additional capital cost is $27 million.

Technical issues identified for emphasis by the Japanese include verification of the reliability and long-term behavior of various alloys at high temperatures. Because variations in temperature and pressure, kind and quality of fuel, and flue gas composition may affect reliability, they feel this must be verified on an actual plant. Development requirements include high-temperature boiler tube alloys; two-stage superheat steam temperature control methods; design of turbine main steam valve, rotor, and blade material; and safety, shutoff, and drain valves.

These improvements would be developed in a step-wise manner to ensure reliability. Initial efforts begun in 1981 focus on qualifying materials at 1100°F (593°C) and then advancing to 1150°F (627°C).

Construction of a commercial plant incorporating 4500 psi (31 MPa) and 1050–1100°F (560–593°C) steam conditions is planned to begin by mid-1986 and to be completed in 1990. Beyond this, construction is to begin in 1990 and be completed in 1995 on a plant representing first commercial application of the full 5000 psi (34.5 MPa), 1150–1200°F (627–649°C) steam conditions. Estimated development costs over the period 1981 to 1987 are about $20 million. □

- Maintenance practices. Each generating unit in Japan must be shut down annually to undergo a detailed inspection and maintenance period. This typically lasts for three or four weeks and is quite extensive, with about 100,000 man-hours expended. Maintenance personnel are trained and assigned specific job functions rather than being rotated through the plant.

- Components. The Japanese concluded, for example, that pulverizer reliability was substandard and have undertaken development of fundamentally improved technology. This effort appears to be very successful.

- Environmental control. The Japanese view environmental control as an integral part of their power plants and lay great emphasis on designing these plants as combined combustion and environmental control systems. As a result, a high level of emission control has been accomplished while minimizing the impact on plant performance and reliability. They are also required to meet stringent NO$_x$ control performance and have implemented an intensive development effort resulting in burner designs that produce half the NO$_x$ emissions typical for a new plant in the United States, while maintaining combustion efficiency.

- Improved reliability. Reliability levels for Matsushima and other coal-fired plants planned by the Japanese, while all include extensive particulate, SO$_2$, and NO$_x$ control, are above the U.S. industry average—but comparable to that of the best U.S. facilities with less auxiliary environmental control. This has been accomplished by dedicating maintenance capability, by maintaining adequate margins in design and operation, and by focusing the necessary development work on critical components and auxiliaries.

In an effort to further assess and evaluate advances the Japanese are making in these areas, EPRI is now organizing a mission of utility industry delegates from across the country to examine Japanese power plant practices in detail. It is anticipated that approximately 15 individuals will make up the group, 3 from EPRI's Coal Combustion Systems Division and 12 from diverse utilities, with expertise in such areas as environmental control, plant design and engineering, plant operation, and financing.

Once the mission is assembled, six study groups will be formed, one each to look at boilers, turbine generators, air pollution control, water pollution control and solid-waste management, auxiliary equipment, and overall plant design. The primary objective is to produce an in-depth analysis of Japanese power plant technology and practice, including a thorough assessment of engineering and construction practices, costs, thermal and environmental performance, and unit availability. This information will then be documented and interpreted by the delegates and EPRI staff, and subsequently published by EPRI.

"Many of the changes the Japanese have incorporated, many of the advances they've made, and many of the practices they've adopted are the result of attitudes we in the United States would do well to consider," comments Yeager. "Primary among these is a willingness to constantly and critically examine our practices and procedures on a value basis and not on the basis of tradition. Beyond this, closer cooperation between utilities, equipment and plant designers and manufacturers, and government would be an important step.

"Because of the differences between our societies, our economic situations, energy mix and needs, and the like," Yeager explains, "adopting this philosophy wouldn't necessarily mean we would wind up making the same decisions as the Japanese. But we should recognize that if we do not make some adjustments, some changes, our preeminence in coal utilization in both domestic and world markets is threatened. In that event, our only likely option to maintain the same levels of performance 5 to 10 years hence would be to rely on foreign technology."

The strategy Yeager recommends to maintain the U.S. edge in technology is to use the present period when little new capacity is being committed here to step up our efforts to monitor Japanese technical development programs and power plant experience; to implement selected parts of the Japanese approach to power plant design and operation, taking a hard-nosed attitude on what will work best for us as opposed to incremental changes in the way we've always done it; and to pursue an aggressive U.S. development program on those advances having merit for both domestic and international conditions.

An important effort is already under way within the utility industry through EPRI's work to characterize the potential for advancing pulverized-coal-fired plant performance. This will include validation of and designs for both new and retrofit applications. Moreover, an 18-month, EPRI-sponsored planning program will soon begin to consider scheduling, development tasks, and costs to validate these power plant technology improvements and develop application guidelines with the utility industry. For these efforts to be successful, however, Yeager underscores that the strong commitment and joint participation of U.S. power plant manufacturers and suppliers are required.

"EPRI looks forward to working with U.S. manufacturers and suppliers to maintain the technical leadership needed for improving pulverized-coal-fired performance and reliability," he says. "Clearly, joint user-supplier efforts in this area will be required to make it possible for U.S. utilities to continue to look domestically for the technology they need to maintain the most efficient and reliable operation of conventional coal-fired power stations and for U.S. suppliers to compete successfully in the world market."

This article was written by William Nesbit, science writer. Technical background information was provided by Kurt Yeager, director, Coal Combustion Systems Division.

Policy options for controlling sulfur dioxides in the United Kingdom are examined. The article categorizes control strategies as pre-combustion, combustion and post-combustion and acts as a good overview of the various systems available. The table on p. 94 is a very good summary of various technologies, with estimates of when they will become available.

Sulphur dioxide from electricity generation

Policy options for pollution control

N.H. Highton and Michael G. Webb

The authors examine the policy options which are available for the control of sulphur dioxide emissions from conventional thermal power stations. Knowledge of these options, their costs of implementation, their probable effectiveness, and the dates when they can be implemented, is necessary for the taking of rational decisions on, for example, the specification of optimal SO_2 emission standards. The emphasis is on the technical aspects of the alternative control technologies. The effects of SO_2 emissions on health and the environment are considered briefly and trends in SO_2 concentration levels and emissions are discussed. Policy options for the control of SO_2 emissions from conventional thermal power stations are then reviewed and assessed.

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The generation of electricity from conventional thermal power stations is associated with a number of environmental problems. For example, generation using coal causes significant air pollution due to emissions of carbon dioxide, sulphur oxides, particulates and nitrogen oxides. In the UK a 2 000 MW coal fired station operating at 60% load factor burns about 4.4 million tonnes of coal per year and each year emits into the atmosphere about 10 million tonnes of carbon dioxide, 130 000 tonnes of sulphur dioxide, 40 000 tonnes of nitrogen oxides, and between 4 000 tonnes and 40 000 tonnes of particulate matter depending on how well the stack emissions are cleaned before they are released.[1] A similar station in the USA, assuming coal seams 10 ft thick, would also require the strip mining of approximately 200 acres each year.

Particular concern has been expressed about the emissions of sulphur dioxide. There are a number of reasons for this. One is that the use of tall stacks to disperse emissions can lead to problems of transnational pollution. A well known example which has received attention in recent years concerns the airborne pollution which has been falling as diluted sulphuric acid over Northern Europe, some of which originates in the UK and West Germany.[2] Another reason is that it has been forecast that UK demand for coal might increase from 121 million tons in 1976 to between 135 million tons and 200 million tons in 2000.[3] The same data source estimates that the power station demand for coal in 2000 might be in the range 75-95 million tons compared with 75 million tons in 1976. Taking the higher of these two estimates the increased coal burn could, in the absence of appropriate control measures, lead to a significant increase in the relative proportion of total SO_2 emissions coming from power stations. In 1976 approximately 50% of all SO_2 emissions in the UK were from power stations. While total SO_2 emissions have been falling in recent years the relative importance of emissions from power stations has been increasing.[4]

Risks to health and the environment

The intention here is not to provide a basis for the estimation of damage cost functions[5] but to draw attention to some of the problems involved in the precise identification of the effects.

To some extent the colourless gas SO_2 is a natural constituent of the air and is, for example, produced by the decay of vegetation. In industrialized countries, however, only about one-tenth of total emissions are estimated to be from natural sources.[6] Emissions of SO_2 from man made sources are associated with the combustion of fossil fuels. Roughly speaking, the burning of 1 kg of sulphur produces 2 kg of sulphur dioxide and sulphur is present, in varying quantities, in both coal and oil.

There are a number of respects in which SO_2 pollution represents a potential health hazard. A WHO committee[7] has pointed out that concentrations of smoke and sulphur dioxide simultaneously exceeding 100μ g/m^3 for long periods can produce respiratory symptoms.[8] Children exposed to these concentrations may have lower levels of pulmonary function and a higher incidence of respiratory disease. Concentrations of smoke and sulphur dioxide exceeding 500μ g/m^3 for 24 hours have been accompanied by excess mortality, especially amongst those already suffering from cardiac or pulmonary disease.[9]

It is important to understand, however, that these measurements are not precise and that it is difficult to distinguish between the effects of the various pollutants.[10] This is a general problem since gaseous and particulate ambient pollutants are rarely found in isolation. Furthermore, there is little understanding of the effects of long-term exposure to small concentrations.

In a more general environmental context air pollution has a variety of effects on man made materials, eg corrosion of metals, and deterioration of fabrics and paints.[11] Although there are suggestions that SO_2 is one of the more offensive pollutants, the specific effects of SO_2 and associated acid rain are very difficult to measure. It is known that smaller SO_2 concentrations than those which cause annoyance to man can damage sensitive vegetation. However, in practice it is difficult to assess the effect because the biological response of a plant to air pollution is a complex function of biological, environmental and climatic factors. Furthermore the concentrations of SO_2 which are typically encountered are rather less than those which have been studied for their effects on vegetation. In general little is known about the effect on trees although their growth may be adversely affected.

The effects of acid rain and the resulting acidification of soil are not well understood, although in general it seems likely that plant growth is inhibited. However, sulphur is also an essential plant nutrient and thus sulphur deficit soils could benefit. But in the UK such soils are rare.[12]

During the past decade, associations between the increased acidity of rain and the acidification of surface waters have been reported from North America and Norway. At present, however, the role of atmospheric sulphur and acid deposition in the contamination of freshwaters is not well understood. The composition of river and lake water is affected by leaching and surface runoff which in turn is affected by the composition of the soil and by the deposition of other pollutants on land.

Although SO_2 emissions can have various adverse consequences, the calculation of reliable and accurate damage cost functions poses many problems of causation and of measurement. These problems are aggravated by the complex relationship between emissions and

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[1] This coal would have an average sulphur content of 1.6%.

[2] *The OECD Programme on Long Range Transport of Pollutants, Measurements and Findings*, OECD, Paris, 1977.

[3] *Coal For The Future*, Department of Energy, London, 1977.

[4] *Clean Air*, Vol 7, No 27, Winter 1977, p 5.

[5] *Economic Measurement of Environmental Damage*, OECD, Paris, 1976.

[6] Health and Safety Commission, *The Hazards of Conventional Sources of Energy*, London, HMSO, 1978, p 13.

[7] World Health Organization expert committee, *Air Quality Criteria and Guides for Urban Air Pollutants*, Technical Report Series, No 506, Geneva, 1972.

[8] Sulphur dioxide concentrations are typically measured in terms of micrograms/m^3 or alternatively in parts per million. The conversion is 1 ppm = 2 860 μg/m^3.

[9] WHO expert committee, *op cit*, Ref 7, p 17.

[10] See L.B. Lave and E.P. Seskin, *Air Pollution and Human Health*, Resources for the Future, Johns Hopkins University Press, Baltimore, MD, 1977.

[11] T.E. Waddell, *The Economic Damages of Air Pollution*, EPA – 600/5 – 74 – 012, US Environmental Protection Agency, Washington, DC, 1974.

[12] Department of the Environment, *Effects of Airborne Sulphur Compounds on Forests and Freshwaters*, Pollution Paper No 7, London, HMSO, 1976, p 12.

concentrations at specific sites. The pollution characteristics of any area are very dependent not only on emissions, but on how they are transported and diffused into the atmosphere. This in turn depends on meteorological conditions. It is also necessary to consider at what height the SO_2 is emitted, since, in general, the higher this is the greater the tendency for the pollutant to be dispersed over a wide area. A recent study of the Forth Valley in Scotland has confirmed the importance of these various factors.[13]

This study employed a numerical dispersion model to quantify the contribution made by the various sources to observed concentrations in the area. In essence this type of model simulates the diffusion of pollutants in terms of meteorological conditions and can be used to predict the impact of future changes in the emission pattern. In this case wind speed and temperature were found to be the most important factors in explaining daily variations, although the effect of wind was dependent on the height of the emissions.

For ground level sources the effect of increasing wind speed was to increase dispersion and thus reduce concentration in the immediate vicinity. In the case of high level sources wind speeds of up to 3-5 m/s brought pollution to the ground more quickly and local concentrations were increased. Increases in wind speed above this level tended to increase dispersion from tall stacks.

The highest ground level concentrations tended to be found in areas where emissions from low and medium level sources were greatest and at those times when there were low level temperature inversions and low wind speeds. Typical winter SO_2 levels between 1973 and 1976 were 32 $\mu g/m^3$ in rural areas and 47 $\mu g/m^3$ in urban areas, these levels being comparable with those reported for other industrial regions of the UK.[14]

The effectiveness of tall stacks in dispersing SO_2 emissions has been confirmed by surveys around six new 2 000 MW stations in the UK. Measuring instruments were placed strategically around the sites and concentrations were compared before and after start-up. Despite considerable variation in local climatic and topographical conditions between the sites, none of the experiments showed any evidence of the effect of a major new power station in the vicinity.[15]

For modern power stations of this type SO_2 concentrations reach a maximum at a distance of between 5 and 20 km from the station. By the time the plume has travelled about 100 km the gases become fairly evenly dispersed in the atmosphere. If it is raining 'washout' or 'rainout' will prevent most of the effluent from travelling further than this. However, in dry weather removal of SO_2 is restricted to slow absorption at ground level by plants, soil and water. Long range transport of SO_2 can then occur with deposition ultimately taking place outside national boundaries.

Trends in SO_2 concentration levels and emissions

The formulation of optimal control policies will require data on the trends of, and forecasts for, SO_2 concentration levels and emissions. There is evidence that despite increases in energy consumption there have been reductions in SO_2 concentration levels in many urban areas. This was partly due to the use of cleaner fuels in response to changes in relative prices and partly due to local environmental legislation.[16]

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[13] A.W.C. Keddie *et al*, *The Measurement, Assessment and Prediction of Air Pollution in the Forth Valley of Scotland*, final report, Warren Spring Laboratory, Stevenage, UK, 1978.

[14] Low and medium level emissions are principally from chimneys less than 100 metres high. Low level emission is mainly associated with the domestic use of coal and oil, as well as that of commercial and public services. Industrial sources account for medium level emissions. High level emissions are generally associated with power stations, although refineries make a small contribution in this category.

[15] A.J. Clarke *et al*, 'Routine sulphur dioxide surveys around large modern power stations', *Atmospheric Environment*, Vol 10, No 4, 1976.

[16] *Environment and Energy Use in Urban Areas*, OECD, Paris, 1978, p 28.

Table 1. Urban SO_2 concentrations (average annual values in $\mu g/m^3$ (rounded to nearest $5 \mu g/m^3$)).

	1964	1965	1966	1967	1968	1969	1970	1971	1972	1973	1974	1975
New York City[a]	–	–	–	–	–	–	220	115	70	75	60	70
Tokyo City	–	225	235	190	180	155	120	100	65	80	70	–
UK[b]	130	115	105	108	110	105	100	95	90	85	80	–

[a] Refers to first quarter for 1973, 1974 and 1975.

[b] Refers to mean urban concentrations – the figures refer to financial years.

Sources: D.L.R. Bailey, C.F. Barrett and I.C. Cooper, *Levels of Smoke and Sulphur Dioxide at Remote Sites in the United Kingdom, 1969-1975*, Warren Spring Laboratory, Stevenage, UK, 1978, Figure 20. *Environment and Energy Use in Urban Areas*, OECD, Paris, 1978, p 31 and p 36.

In New York City, for example, there was a considerable decline in SO_2 concentrations in the late 1960s, a trend which continued into the 1970s. This was accompanied by rising energy consumption combined with specific measures to control emissions of both sulphur oxides and particulates from stationary sources. Legislation to control air pollution in Tokyo was passed in 1967, 1968 and 1969. This was aimed at reducing the sulphur content of heavy fuel oils and converting heat sources in newly constructed buildings, as well as a variety of other measures. Recently a clear downward trend in SO_2 levels has been observed. Annual average concentrations recorded in Tokyo city in 1974 were less than half those recorded in the latter half of the 1960s. The fuel control measures were thought to be mainly responsible for these improvements.

Recorded concentrations in New York City, Tokyo City and UK urban areas are presented in Table 1.

In the UK at the national level there has been no great change in total SO_2 emission over the past decade, although a downward trend since 1973 is discernible. There has, however, been a clear change in

Table 2. Estimated SO^2 emissions from fuel combustion in the UK, 1969-76 (million tonnes).

Source	1969	1970	1971	1972	1973	1974	1975	1976
High level								
Power stations								
Coal	2.12	2.13	2.01	1.84	2.07	1.81	2.15	2.10
Gas oil	0.01	0.01	0.01	0.02	0.01	0.01	0.00	0.00
Fuel oil	0.44	0.63	0.79	1.01	0.94	0.96	0.80	0.67
Total	2.57	2.77	2.80	2.87	3.02	2.78	2.96	2.78
Refineries								
Fuel oil/gases	0.24	0.24	0.25	0.26	0.29	0.30	0.26	0.29
Total high level	2.81	3.01	3.05	3.13	3.31	3.08	3.22	3.06
Medium level								
Coal	0.68	0.60	0.49	0.36	0.36	0.33	0.27	0.25
Fuel oil	1.17	1.19	1.14	1.09	1.08	0.99	0.90	0.81
Other	0.28	0.28	0.27	0.29	0.29	0.25	0.25	0.25
Total medium level	2.13	2.08	1.90	1.72	1.73	1.56	1.42	1.30
Low level								
Coal	0.51	0.47	0.41	0.33	0.31	0.29	0.24	0.23
Fuel oil	0.25	0.26	0.24	0.20	0.17	0.16	0.15	0.13
Other	0.29	0.27	0.25	0.27	0.26	0.24	0.23	0.23
Total low level	1.08	1.02	0.90	0.80	0.76	0.72	0.64	0.61
Grand total	6.02	6.12	5.86	5.65	5.81	5.36	5.28	4.98

Note: Totals include rounding errors and small contributions from sources not included in the table.

Source: M-L Weatherley, *Fuel Consumption, and Smoke and Sulphur Dioxide, in the United Kingdom, up to 1976*, Warren Spring Laboratory, Stevenage, UK, 1977, p 6.

Table 3. Estimated SO_2 emissions from coal combustion in the UK up to 2000 (million tonnes).[a]

Source[b]	1970/1	1975/76	1980	1985	1990	2000
High level	1.78	1.96	2.14	2.43	2.40	2.35
Medium level	0.69	0.31	0.31	0.31	0.31	0.31
Low level	0.44	0.32	0.21	0.18	0.18	0.18
Total	2.91	2.59	2.66	2.92	2.89	2.84

[a] Includes all solid fuel usage. Figures rounded to two decimal places.

[b] All power station emissions are taken as high level, all industrial emissions as medium level and all domestic emissions as low level.

Source: National Coal Board, private communication.

the relative importance of emissions from high level sources. Table 2 shows how the relative contribution from power stations has increased.

It might be expected that an increase in the proportion of SO_2 emissions from high stacks would be accompanied by a decline in local concentrations and a slight increase in concentrations at remote sites. The data given in Table 1 confirm that, in the UK, urban SO_2 levels did decline over this period. However, measurements made at a number of remote sites between 1969 and 1975 showed no discernable increase. The mean annual concentration in eastern England over this period was 14-28 $\mu g/m^3$. In the west the concentration was 7-10 $\mu g/m^3$, in southern Scotland 10 $\mu g/m^3$ and at the most northerly site near Lerwick it was 1.5 $\mu g/m^3$.[17] The trend in urban concentrations when combined with fairly constant levels at rural sites suggests that tall stacks are highly effective in dispersing pollutants.

Decisions on control policies require estimates of both the total level and the future pattern of SO_2 emissions. Estimates for the UK to the end of this century are given in Table 3. Only coal combustion is included although, as was shown in Table 2, this is the largest single source of emissions. These projections take into account National Coal Board production plans and possible future changes in the sulphur content of coal.[18]

Broadly speaking these projections indicate a continuation of the trends established over the last decade. While no great change is forecast in total emissions an increase is expected in the proportion of high level emissions. For comparative purposes forecasts for Canada, Japan, USA, Japan and OECD Europe are given in Table 4.

To sum up, emissions from power stations in the UK have become relatively more important and this trend is expected to continue in the future. Also, high level sources do not affect local concentrations, but they do contribute to transnational air pollution.

Policy options for the abatement of SO_2 emissions

Broadly speaking the available abatement policies[19] fall into three categories:

- Emissions may be controlled before combustion by reducing the sulphur content of fossil fuels burned. This could involve the use of coal and fuel oil desulphurization technologies or the use of fuels naturally low in sulphur.
- Emissions may be controlled during combustion by employing modified combustion techniques.
- Emissions may be controlled after combustion by employing flue gas desulphurization processes.

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[17] D.L.R. Bailey, C.F. Barrett, and I.C. Cooper, *Levels of Smoke and Sulphur Dioxide at Remote Sites in the UK, 1969-1975*, Warren Spring Laboratory, Stevenage, UK, 1978.

[18] Note that the figures in Table 3 have not necessarily been calculated on the same basis as those in Table 2.

[19] The policies considered are those which would be available if emission standards were imposed on power stations in the UK. Most of the options are nevertheless applicable to other countries. Cost considerations are outside the scope of this paper. They are, however, part of our continuing work and will be reported in a later paper.

Table 4. Estimated SO_2 emissions in Canada, Japan, USA and OECD Europe in 1974 and projections for 1985 (million tonnes)[a]

	1974	1985[b]
Canada	1.42	1.67 - 1.84
Japan	2.44	2.55 - 3.64
USA	22.75	23.18 - 26.28
OECD Europe	19.72	22.05 - 25.42

[a] Restricted to fuel combustion in stationary sources.

[b] These figures assume a continuation of present policy governing supply expansion and conservation. The ranges reflect different sources of crude oil supply and variations in desulphurization capacity.

Source: *Clean Fuel Supply*, OECD, Paris, 1978, pp 18-28.

Although some of these technologies are well established others are not expected to become commercially available for many years. An additional policy, and one which could have effect in the short run, would be to modify the operation of existing power stations. Essentially this would involve changing the 'merit order' in which plant is brought into operation. The aim would be more intensive operation of power stations with low emissions of SO_2 per kWh generated.

Changes in merit order

Power stations vary in the amount of SO_2 emitted per kWh generated. For example it is relatively high for a station with low thermal efficiency burning high sulphur content coal. At the other extreme, there are no SO_2 emissions from nuclear stations.

The following simple principle would be used in a policy to reduce the aggregate SO_2 emission from power stations by changing the merit order. Total SO_2 emissions would be reduced if a power station with high SO_2 per kWh changes place with another one lower down the merit order which has low SO_2 emission per kWh. These substitutions of output should continue until all opportunities to reduce SO_2 emissions have been exhausted. At this point the aggregate SO_2 emission from existing plant would be minimized. If the higher costs associated with these plant substitutions were passed on in the form of higher prices then, if demand is price elastic, there will be a secondary reduction in SO_2 emissions. It will be convenient to assume that the price elasticity is zero so that this effect can be ignored.

In fact this policy provides negligible scope for a reduction of emissions in the UK. This is because the existing merit order is already optimal with respect to SO_2 abatement. That is, the existing merit order operation would remain largely unchanged if the CEGB was suddenly required to pay for each kg of SO_2 emitted. This is to be expected since nuclear stations have the lowest operating costs and, in fossil fuelled stations, thermal efficiency is dominant in determining both SO_2 emission and operating cost per kWh.

Similar proposals have been put forward for temporary control of emissions where meteorological factors cause undesirable dispersion. Basically this is aimed at reducing the offensiveness of SO_2 emissions by shifting the burden of generation from one area to another in order to integrate with local weather conditions. It is not aimed at reducing total emission, in fact it could increase it. It is believed that the potential advantages of this policy are small. Furthermore the departure from strict merit order operation could increase costs substantially. However, policies of this type do have an advantage in that they can be implemented in the short term.

Control before combustion

This option effectively encompasses three possible policies: the increased use of low sulphur oil; the desulphurization of oil; or the desulphurization of coal.

Use of low sulphur oil. Although the natural sulphur content of oil can be regarded as fixed, the average content of all crude oil used in the UK can be reduced by changing the pattern of oil imports and exports. Thus North Sea crude oil is relatively light and low in sulphur content and if its use was mainly confined to Europe there would be some scope for a reduction in the sulphur content of fuel oil in Europe as a whole. It has been forecast that under these circumstances the average sulphur content of fuel oil in Europe could be in the range 1.7% to 2.1% by 1985 compared with an estimated average content of 2.5% in 1974.[20] If, alternatively, North Sea oil was mainly used outside Europe or was used to displace imports of similar crudes then average sulphur content might be in the range 2.3% to 2.7%. Alterations in the pattern of supply of crude oil would however have complex ramifications on the market for oil products as a whole and the oil refining industry. These would have to be fully evaluated.

The implementation of this policy would not lead to a major reduction in SO_2 emissions from power stations since their use of oil is not a major contribution. The potential reductions would, however, be greater if the UK pursued a low sulphur oil policy by itself. The cost of such a policy depends on the price differential between low and high sulphur oils. In the long run this differential is limited by the cost of oil desulphurization.

Oil desulphurization. The desulphurization of gas oil is already widespread. Usually about 90% of the sulphur can be removed with fuel losses of around 3.5%. However, although the technology is proven, fuel oil desulphurization is not standard practice. When used it is incorporated into the refining process and sulphur is obtained as a by-product. The techniques are normally classified into direct and indirect desulphurization. The direct process involves treatment with hydrogen at high temperature and pressure in the presence of a catalyst, and it can remove up to 90% of the sulphur. This process is technically more difficult than the indirect process and does not work well with residues high in asphaltenes and metals.[21] Capital investment costs are relatively high, and they are increased significantly if the process is retrofitted to an existing refinery. No technical developments appear to be anticipated which would significantly affect these costs. Typically the direct process involves about an 8% fuel loss.

Indirect desulphurization involves the vacuum distillation of atmospheric residues. The vacuum distillate is desulphurized in a process similar to conventional gas oil desulphurization and is blended back with the vacuum residue.[22] This allows the desulphurization of crudes which are difficult to treat directly. Since only a proportion of the fuel oil is effectively treated the sulphur content is reduced by only 30-40% with an associated fuel loss of about 5%. The capital and operating costs are lower than those of the direct process.

It is important to note that it is thought impossible to reduce the

[20] These figures apply to the European Countries of OECD. See *Clean Fuel Supply*, OECD, Paris, 1978, pp 35-37.

[21] Chemical Systems International, *Reducing Pollution from Energy Transformation Sources*, Graham and Trotman for the EEC, London, 1976, p 51.

[22] G.A. Persson, 'Control of sulphur dioxide emissions in Europe', *Ambio*, Vol V, Number 5-6, 1976.

sulphur content of UK fuel oil to 1% before about 1990 because of constraints on the suppliers of capital equipment.

Coal desulphurization. There are a large number of methods of physically cleaning coal which are mainly designed for removing dirt and for separating the coal into homogeneous sizes to meet market requirements. However, they are also capable of removing some sulphur. The most established techniques involve washing finely ground coal and in 1974 87% of UK coal was cleaned.[23]

Sulphur in coal is present in two forms, pyritic and organic. Although the proportions vary with different coals, high sulphur content tends to be accompanied by relatively high pyrite content. Pyrites are usually in the form of fine particles which means that techniques for the treatment of fairly small coals can be successful in removing them. Unfortunately organic sulphur cannot be removed by washing and thus high sulphur coals tend to be best suited to this treatment because of their relatively high pyrite content.

Various techniques are available for both coal washing and for the physical removal of pyrites. In the UK coal washing is principally done using hydraulic jigs and dense medium baths. Ground coal is suspended in a fluid, the refuse material falls to the bottom of the separating unit and the clean coal moves to the top for removal. This method is not very efficient for removing pyrites. An alternative technique appropriate for fine coal uses froth flotation and is based on the different surface properties of the different particles. Unfortunately pyrites have similar surface properties to coal and thus in normal operation this technique is not efficient in separating out the sulphur. However, there is some removal because of density differences.

Among the physical methods available for pyrite removal the most widely used is probably the centrifugal dense medium technique. This is relatively efficient and can give two product streams with the pyrites concentrated in one of them.

In general the suitability and effectiveness of these techniques depends heavily on the nature of the coal being treated. In general sulphur removal is easier if the pyrite particles are relatively large. If the particles are small then the avoidance of large losses of coal requires that it be crushed into correspondingly smaller sizes, with a consequent increase in costs.

As a guide, between 35% and 70% of pyritic sulphur can be removed by physical washing.[24] However, the organic sulphur which is not removed by this process may make up between 30% and 70% of the total sulphur content. This suggests that, overall, only about 25% to 30% of the sulphur will be removed. For illustration about 16% of total sulphur content is removed from coal from the Durham (UK) field by physical washing, but it should be noted that no attempt is made to optimize the operation for sulphur removal.

As a method of controlling SO_2 emissions this option is relatively cheap, although the total abatement potential is not very great. Reliability should not be a problem since the techniques are well established. Available evidence suggests that it is also possible to dispose satisfactorily of any waste which is a potential hazard to the environment.

A number of processes have been suggested for cleaning coal by chemical methods, but most of these have only been proved on a laboratory scale. They tend to be efficient in removing pyritic sulphur, have low thermal losses, relatively high costs, and to be associated with various environmental problems. At the present time the most developed of these processes is the TRW Meyers Process. This is 95% efficient in removing pyritic sulphur but has byproducts of sulphur and a waste material which must be disposed of carefully since it is an environmental hazard. The US Environmental Protection Agency (EPA) has suggested that the process could be commercial in five to six years time. On a similar time scale the EPA has suggested that another method, known as the Battelle Hydrothermal Process, could become commercial. It is estimated that this could remove up to 99% of pyritic sulphur and about 25% to 70% of organic sulphur, depending on the type of coal.

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[23] National Coal Board, private communication.

[24] Environmental Protection Agency, *Electric Utility Steam Generating Units, Background Information For Proposed SO_2 Emission Standards*, EPA – 450/2-78–007a, US Environmental Protection Agency, Washington, DC, July 1978, pp 4-9.

Another potential method is the Solvent Refined Coal (SRC) process which can be used in conventional boilers. Pulverized coal is dissolved in a coal-derived solvent at high temperature and pressure. Small quantities of gas and light liquids are produced along with a heavy organic material which is SRC. The process removes nearly all the ash and up to 85% of pyritic and organic sulphur. Energy losses, however, may be relatively high at about 30%. Because of serious technical difficulties which remain to be overcome it is believed that this process will not be commercial until the mid 1990s.[25] The main environmental problem is the centralization of ash and sulphur disposal. This occurs because in commercial operation plants would be large and near coal mines. In effect the waste disposal problems of several power stations would be localized.

Techniques are also being developed for the gasification of coal for use in conventional boilers. Sulphur removal is expected to be about 90%, energy losses about 30%, and costs are estimated to be relatively high. Although the process may become commercial in the 1990s it is not expected that gasified coal would be economical for electricity generation in the UK. At this stage a somewhat speculative idea is to gasify coal for use in advanced power cycles. This is a complex process in which generation and gasification are integrated. Thermal efficiency is potentially high with low SO_2 emission. The technology is, however, highly uncertain and it is not expected to be commercially feasible until the mid-1990s.[26]

Control during combustion

The main interest in this area has been fluidized bed combustion (FBC). This has a number of potential advantages over conventional boilers. These include a high heat transfer rate and the possibility of reducing emissions of both SO_2 and nitrogen oxides.

In FBC solid particles of coal, ash and sorbent material are held in suspension by air injected at controlled rates. Combustion, at either atmospheric or higher pressure, takes place in the fluidized bed and, typically, steam is produced in tubes or water walls located in this region. Pressurized fluidized bed combustion (PFBC) allows greater combustion efficiency and smaller size. However, atmospheric pressure fluidized bed combustion (AFBC) is both simpler and more developed than PFBC.

Sulphur dioxide absorption is achieved in AFBC with the injection of limestone. The limestone requirement is approximately 15% by weight of the coal input. It does, however, vary widely with the sulphur content of the coal and the desired absorption efficiency. In

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[25] Environmental Protection Agency, *Energy – Environmental Considerations of Flue Gas Desulphurisation*, US Environmental Protection Agency, Washington, DC. Two SRC pilot plants are currently operating in the USA.

[26] *Ibid.*

PFBC dolomite is a more efficient sorbent than limestone although at least an extra 60% (by weight) is required. Removal efficiency in AFBC systems using limestone can be up to about 90% depending on the quantity of sorbent used. Maximum possible removal in PFBC systems using dolomite is thought to be slightly higher than this.

Byproducts from FBC consist of ash and spent sorbent. This is in the form of handleable dry granules. Since the waste is not believed to have any toxic effects it should be possible to use it as a landfill with a minimum of preparation. The main disposal problem arises from the large quantities which are involved. The additional waste arising from SO_2 removal is roughly comparable with the normal volume of coal ash.

The large volume of sorbent and waste involved with SO_2 removal in FBC has led to research into methods of regenerating the spent sorbent and recovering sulphur or sulphuric acid. So far no satisfactory method has been devised, and available evidence suggests that the energy costs of recovery and regeneration would probably be high.

The costs of both AFBC and PFBC are uncertain, although some estimates show savings over a conventional system.[27] Capital costs are not greatly affected by SO_2 removal, although operating costs are increased in a number of ways. The raw material costs of limestone and dolomite are generally low, although these can be increased by local shortages and high transport costs. Waste disposal costs vary considerably and depend mainly on whether this is done on- or off-site. The disposal costs associated with PFBC are appreciably higher than those of AFBC due to the additional volume of waste.

The overall conversion efficiency of FBC is high. In the case of PFBC this can be further enhanced in steam and gas turbine systems. SO_2 removal does, however, result in some loss of energy. This arises from extra material preparation and feeding as well as the removal of hot solids.

There are currently no power stations employing FBC and all available information derives from small scale test plants. It is possible that problems will be encountered when these are scaled up. Although FBC is regarded by the EPA as a theoretically satisfactory method of SO_2 control, whether it will be economical and practicable depends on the course of research and development. Estimates of the date for the commercial availability of AFBC range from 1985 to 1990, and those for PFBC lag a few years beyond.

Control after combustion

At least twenty different flue gas desulphurization (FGD) processes have been or are being tested. Broadly speaking they fall into two categories. The processes are either throwaway, in the sense that by-products are useless, or regenerable in the sense that sorbent material is recycled and sulphur or sulphuric acid is recovered as a potentially marketable product.

They can also be classified into dry or wet. Wet scrubbing processes involve the contact of flue gases containing SO_2 with slurries or solutions. So far, dry processes have not been regarded as successful and have generally received less attention. One of the advantages of wet scrubbing is that fly ash as well as SO_2 can be removed, although in some systems pre-removal of fly ash is desirable for efficient SO_2 removal. A disadvantage is that the use of slurry

[27] Environmental Protection Agency, *op cit*, Ref 24.

tends to result in scaling, plugging and the erosion of equipment, although this is minimized with the use of solutions.

All wet processes suffer from the disadvantage that treated flue gases experience considerable cooling. Normally flue gases are discharged at about 120°C but wet scrubbing reduces the temperature to about 50°C. Unless the gas is reheated water vapour and sulphur oxides can condense and corrode stack linings and other equipment. The rise of the plume is also impaired, resulting in poorer dispersion of residual pollutants for a given stack height. Water vapour can condense which makes the plume visible and droplets can settle around the station. Another disadvantage of wet processes is that waste products tend to be more difficult to dispose of.

A number of general remarks can be made about the major determinants of SO_2 removal in wet scrubbing systems:

- SO_2 absorption is improved if the volume of scrubbing liquid is large compared with the volume of flue gas. This allows concentrations of dissolved SO_2 to be kept low and permits a large gas–liquid interface area. On the other hand pumping requirements are increased.
- Low gas velocity increases residence time in the scrubbers and increases absorption. This is also aided by having a relatively long residence time of the scrubbing medium in the scrubber.
- The scrubber design is highly important. Several types exist and the main aim is to maximize the gas–liquid interface area. One design consists of a tower filled with mobile packing spheres supported by grids. Flue gases pass up through the tower counter-current to the scrubbing liquid which is fed in from above. Another type of scrubber is the spray tower into which the absorbent is sprayed through an array of nozzles. Various other designs have been tried.

Energy losses in FGD arise mainly from three causes; the circulation of the scrubbing medium, the fans which compensate for loss of pressure across the scrubbing device, and the reheat of flue gases. Energy requirements tend to be an increasing function of SO_2 removal. Reheat can be reduced by mixing some untreated gas with scrubbed gas, but this reduces the overall absorption efficiency.

In the case of regenerable processes a considerable amount of energy is required to recover sulphur and sulphuric acid. In the throwaway processes energy is required to handle and transport the waste. The costs of this are sensitive to whether this occurs on- or off-site. Other energy requirements arise from the pre-removal of particulates which is required with regenerable processes, and with the preparation and feeding of raw material inputs.

In general the energy requirements, when measured as a percentage of electricity generated, are not sensitive to power station capacity although they are sensitive to the sulphur content of fuel. On the whole regenerable processes are more energy consuming than throwaway processes.

Four of the best proven FGD scrubbing systems are lime/limestone, dual alkali, magnesium oxide, and sodium sulphite. The last two are regenerable. All of them are applicable to both coal and oil fired installations and can be retrofitted to existing power stations, although capital costs are higher when retrofitting occurs. We shall now consider some aspects of these four processes, paying particular attention to some of their adverse environmental effects.

Lime/limestone process. The lime or limestone is ground to a fine powder (either on- or off-site) and then mixed with water until the content of solids is 10-20%. Fine grinding reduces the amount of sorbent that remains unreacted. When this slurry comes into contact with the flue gases the SO_2 reacts to produce a waste sludge. Lime is a more effective absorbent than limestone, reducing the volume of liquid required, pumping costs, and also the volume of waste. The main disadvantage is that lime is considerably more expensive than limestone.

This system has a number of advantages. It is relatively simple and cheap with at least 90% SO_2 removal efficiency. Problems of plugging, scaling and erosion have largely been solved with careful scrubber design. The system can also tolerate fly ash. Its major disadvantage is that very large quantities of waste sludge are produced.[28] The chemical and physical characteristics of the sludge vary with a number of factors, for example the coal used, the nature of the scrubber, and the ash collection system. The choice of disposal method will be site specific. The sludge can be ponded on indigenous clay soil or placed in lined pits. Alternatively it can be disposed of at sea or in disused mines. With chemical treatment it may be possible to use if for landfill, although this could be expensive. A problem with this is that 'fixed' waste can revert to sludge when it is frozen and subsequently watered.

The disposal of sludge must be considered very carefully because of possible seepage to groundwater, runoff to streams, and damage to ocean life if it is disposed of at sea. These factors will depend on local climatic and geographical conditions. The disposal sites would have to be carefully monitored since there would be a real danger of trans-media pollution occurring.[29] In order to minimize potential runoff problems it may be necessary to cover the sites with soil and to plant vegetation.

The lime/limestone process cannot be regarded as suitable for areas without a large disposal site, and this will severely limit its possible retrofitting to existing CEGB power stations. For example, a 500 MW plant operating at an average load factor of 50% and aiming at 90% SO_2 removal using limestone absorbent would, depending on the sulphur content of coal, produce between 4 000 and 7 500 acre-feet of waste over a thirty year period. Larger plants would produce additional waste in roughly the same proportion.

This process is fully developed and is ready for application to full scale power stations. Given proper design and maintenance the EPA regard it as an effective and reliable method of sulphur control. There is considerable experience with the process in the USA where at least 24 units are in existence, about half of which were retrofitted. Two examples are the Green River station of Kentucky Utilities and the Kansas City Power and Light La Cygne station.

The 64 MW Green River Station was commissioned in 1949. It is a peak-load unit burning 3.8% sulphur coal. A lime FGD system was retrofitted and started operation in 1975. After some initial start-up problems, mechanical reliability has been good and the design specification of 80% SO_2 removal has been exceeded easily. The 820 MW La Cygne station burns 5-6% sulphur coal and it began operation in 1973 with a limestone FGD unit. Numerous initial problems were encountered, many of which were not associated with the FGD unit. Subsequently operation has been very reliable with 80% SO_2 removal, although the design specification was 76%.

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[28] F. Princiotta *et al*, *Status Report on the Limestone/Sludge Flue Gas Desulphurisation Process*, NATO-CCMS Study, PEDCO Environmental, USA, 1978.

[29] This simply means that the reduction of air pollution by reducing emissions of SO_2 could lead to pollution in another part of the environment, for example in water courses.

Double alkali process. Instead of using a slurry this process involves the scrubbing of the flue gases with a clear alkali solution containing sodium salts. The spent absorbent is then reacted with lime or limestone. The result of this reaction is a regenerated scrubbing solution, which is returned to the system, and a solid waste. This waste product is essentially the same as that produced by the lime/limestone process. In addition to this waste some of the sodium solution oxidizes and must be purged from the system.

The use of a solution rather than a slurry simplifies the scrubber design, thus reducing cost, and also increases reliability. Another advantage is that only a low liquid to gas ratio is required which reduces pumping requirements. However, this two stage process is generally more complex than lime/limestone scrubbing and is more expensive.[30] The SO_2 removal efficiency is in the range 85% to 95%.

There is relatively little experience with this process. Most examples are small coal and oil fired industrial boilers in Japan and the USA, although in Japan there are three operational units in oil fired power stations. Three units on coal fired power stations are being constructed in the USA. In general it is regarded as a viable and successful method of SO_2 control.

Magnesium oxide process. This is a regenerable wet slurry scrubbing process. In contrast to the lime/limestone and double alkali processes there is no sludge to dispose of. The byproduct is marketable and scaling problems are also reduced.

Magnesium oxide scrubbing requires the prior removal of particulate matter from the flue gases before they enter the absorber. The flue gases are brought into contact with a slurry of magnesium oxide and the SO_2 reacts to form magnesium sulphite. This is regenerated into magnesium oxide, which is topped up by about 2% to meet losses during each cycle and returned to the scrubbing system.

The regeneration involves thickening the scrubber effluent from about 10% to 40% of solids. Remaining water is removed by evaporation and by centrifuging. A solid cake emerges which is finally dried by heating. Thus a magnesium sulphite solid is obtained which is chemically stable and may be stored or transported to an off-site regeneration plant. Regeneration back to magnesium oxide and SO_2 is achieved by heating to 1500°F. Concentrated SO_2 can then be used in the conventional production of sulphur or sulphuric acid.

Sulphur and sulphuric acid are basic industrial chemicals with a wide variety of uses in the production of, for example, paper, fertilizers, and in the rubber processing industry. Sulphur is usually obtained from ore deposits. The UK imports about 1.4 million tonnes of sulphur per year compared with a possible output of 0.5 million tonnes from the wide scale adoption of FGD.[31] It is believed that this quantity could be fairly easily absorbed and help reduce imports. Sulphur is relatively easy to store and transport.

In contrast sulphuric acid is expensive to transport and has a more restricted range of industrial uses than sulphur. In 1976 the UK produced about 3.3 million tonnes compared with an estimated annual output from FGD of 1.5 million tonnes. It has been suggested that in the long term the market for sulphuric acid will increase and that FGD acid byproducts could be absorbed with careful marketing and planning.[32] Current market prices would favour the production of acid rather than sulphur.

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[30] G.C. McGlamery *et al. Detailed Cost Estimates for Advanced Effluent Desulphurisation Processes*, EPA 600/2 – 75 – 006, US Environmental Protection Agency, Washington, DC. 1975.

[31] M. Prior, *The Control of Sulphur Oxides Emitted in Coal Combustion*, report No B1/77, International Energy Agency, Coal Research, London, December 1977, p 66.

The potential revenue from marketable byproducts is estimated to be small compared with the other costs of FGD. It would probably do no more than cover the expenses of chemical conversion plants. The main advantage of regenerable processes is that they avoid the production of waste. In areas where waste disposal is a problem this would be an important factor in favour of regenerable processes, and the choice between the processes may be site specific.

Magnesium oxide scrubbing, like other regenerable systems, is a complex process which effectively converts power stations into chemical processing plants for which somewhat different knowledge and expertise is required. A comparative lack of long-term experience of this is a major disadvantage of the system, although it has been tested on three power stations in the USA. One of these is the 155 MW Mystic Station No 6 of the Boston Edison Co which burns 2.5% sulphur fuel oil. This was the first power station in the USA to employ magnesium oxide FGD and although a number of operating difficulties were encountered it is regarded as having proved the technical feasibility of the process. Operating experience of the process is currently being obtained from two other stations, the Dickerson No 3 of the Potomac Electric and Power Co and the Eddystone No 1A station of the Philadelphia Electric Co. These stations burn 2.0% and 2.5% sulphur content coal respectively and about 90% SO_2 reduction has been achieved.

Sodium sulphite (Wellman–Lord) process. This process also regenerates sulphur or sulphuric acid and is a fairly well proven system. Flue gases are cooled and humidified in a water prescrubber, they then pass into the main scrubber where they are washed with a solution of sodium sulphite. This absorbs SO_2 and is converted into sodium bisulphite. This solution is then evaporated by heating, driving off concentrated SO_2 and reforming sodium sulphite crystals. SO_2 is then led to a nearby sulphur or sulphuric acid plant. The main environmental problem of this process is that some of the scrubbing solution oxidizes to produce sodium sulphate. This must be removed from the system and disposed of at a rate of around 1 tonne/hour from a 500 MW station. There may, however, be a limited market for this waste.

Other disadvantages are that fly ash must be kept out of the system and potential corrosion requires the use of expensive materials. It is also associated with relatively high energy consumption. Advantages include a lack of scaling problems and a low liquid to gas ratio requirement.

This process had been successfully operated on two small oil fired industrial boilers in Japan since 1971. In 1973 two larger oil fired systems, including a 220 MW utility boiler, were commissioned in Japan and they have subsequently been operated successfully. Similar experience has been reported from the USA where one power station is currently operating using this process (a 115 MW coal fired unit on the Mitchell station of Northern Indiana Public Service) achieving an

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[32] *Ibid.*

Table 5. Summary of alternative SO_2 control options for power stations.

Control option	Description	Availability	Applicability	Energy losses	SO_2 removal	Environmental problems
Merit order	Operate less polluting plant more intensively	Short term			Negligible under present UK conditions	
Low sulphur crude oil	Substitute North Sea oil for high sulphur crudes	Short term	Oil fired power stations		About 25% (see text)	
Oil desulphurization	Indirect or direct desulphurization incorporated into refining process	Currently available technology	Oil fired power stations	8% (direct) 5% (indirect)	90% (direct) 30%-40% (indirect)	
Coal desulphurization Physical washing	Removal of pyrites. Based on physical differences from the rest of coal	Currently available technology	Coal fired power stations.	5%-20%	10%-50%	Disposal of high sulphur refuse
Coal desulphurization Chemical cleaning	Chemical removal of pyritic sulphur and possibly organic sulphur, depending on technique used	Mid-1980s	Coal fired power stations	10%-25%	40%-80%	Disposal of toxic byproducts
Solvent refined coal	Dissolve pulverized coal. Ash also removed	1990s	Coal fired power stations	30%	85%	Centralization of ash and sulphur disposal
Coal gasification	Sulphuric acid recovered in process. (Efficiency enhanced in advanced power cycles.)	1990s	Gas fired power stations	30%	90%	Undefined
Fluidized bed combustion	SO_2 absorbed during combustion (Efficiency of PEBC enhanced in combined generating cycles.)	1980s (AFBC)	New coal fired power station	See text	70%-95%	Large quantities of spent sorbent
Lime/limestone FGD (throwaway)	Flue gases scrubbed with slurry of lime/limestone	Currently available technology	All fossil fuelled power stations	See text	90%	Very large quantities of waste sludge
Double alkali FGD (throwaway)	Alkali scrubbing solution	Currently available technology	All fossil fuelled power stations	See text	90%	Very large quantities of waste sludge
Magnesium oxide FGD (regenerable)	Magnesium oxide scrubbing slurry	Currently available technology	All fossil fuelled power stations	See text	90%	
Sodium sulphite FGD (regenerable)	Sodium sulphite scrubbing solution	Currently available technology	All fossil fuelled power stations	See text	90%	Disposal of small quantities of sodium sulphate possibly required

SO_2 reduction in excess of 90%. The New Mexico Public Service is currently installing this process on two 700 MW units at its San Jan station, which are due to be commissioned in 1981.

The US EPA regards this process as having proved itself as the most successful regenerative process on oil fired boilers. Its status on coal fired boilers is still being demonstrated.

Conclusions

The possible effect of these control options on the overall level of abatement is complicated by at least two factors:

- The resulting increase in electricity prices could lead to a secondary reduction in emissions. But since the demand for electricity is relatively price inelastic this effect would probably be small.[33]
- Some of the options (eg those associated with the retrofitting of FGD) would reduce a station's thermal efficiency and increase its unit operating costs. For a given level and pattern of demand for electricity there could be an increase in the SO_2 emitted by other stations which increased their output to compensate for that lost at the retrofitted station.

The choice and mix of control options depends in two ways on the chosen abatement standard. First, on the level of abatement which is required and second, on the time at which the standard is introduced. The important factors here are the dates at which the various technologies are expected to become viable and the rate at which investment could be undertaken. The latter is affected by capacity constraints on the suppliers of capital equipment. Also important is the extent to which desulphurization methods can be applied to existing stations rather than confined to new ones. This will be affected by the age structure of existing plant, forecasts of electricity demand, the planned investment programme, and the expected reserve plant margin.[34] The relevant features of the control options which have been discussed in this paper are summarized in Table 5.

This article has been concerned with what may be termed the piecemeal approach to the control of pollution. That is, it was concerned solely with one pollutant. However, at a number of places it was pointed out that the control of SO_2 could lead to a problem of trans-media pollution, in which case a global approach to the control of pollution is really required. The absence of a policy specified on this basis could lead to a situation where the quality of one part of the environment was improved but at the cost of a deterioration of another part. In some ways this is the counterpart of a policy (such as the use of tall stacks) which can lead to a local improvement in the environment at the cost of a deterioration in some more distant environment.

Finally, many of the policy options for the control of SO_2 emissions involve a cost in the form of increased energy consumption. This is simply an example of the trade-off which may be required between an improved environment and energy conservation.

[33] See the evidence in Department of Energy, *Report of the Working Group on Energy Elasticities*, Energy Paper No 17, HMSO, London, 1977.

[34] This will affect any planned retrofitting programme given an adopted standard of service.

This Environmental Protection Agency/Department of Interior report indicates the economic advantages of coal cleaning in combination with FGD, instead of using FGD alone. The cleaned coal burns more efficiently, and uses less coal; all of which helps to reduce the overall cost of electricity production. The readers attention is drawn, in particular, to the summary of case results which demonstrates the economic advantages in selected cases.

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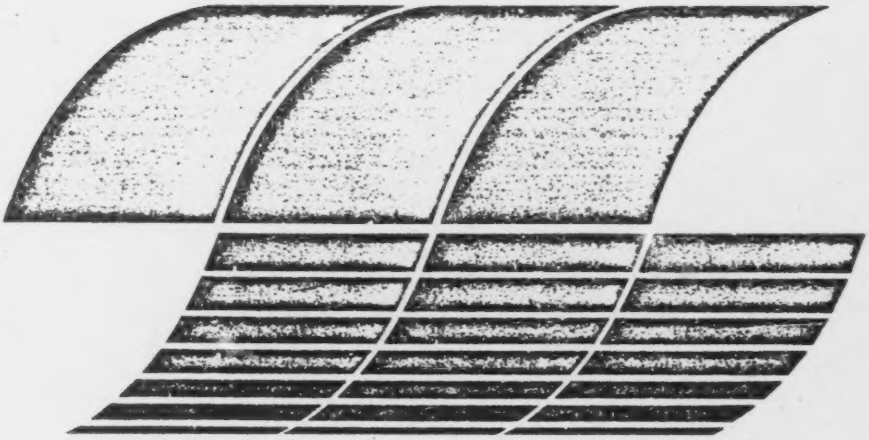
Coal Cleaning with Scrubbing for Sulfur Control:

An Engineering/ Economic Summary



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**Bureau of Mines, U. S. Department of Interior

Coal Cleaning with Scrubbing for Sulfur Control:

An Engineering/ Economic Summary

United States Environmental Protection Agency
Office of Research and Development
Office of Energy, Minerals and Industry

August 1977

Introduction

In this country, the continued generation of power in a reliable, cost-effective, and environmentally acceptable fashion is critical to economic and social stability. Much of this power is produced by coal-fired plants. During 1975, fifty-five percent of the utility industry's fossil fuel requirements were met with coal. In view of the rapidly diminishing reserves and rising prices of alternate fossil fuels, the percentage provided by coal will increase. If these plants are going to adhere to the environmental standards set forth by EPA and others relative to SO_x emissions, it will often be necessary to consume coals of less than 1% sulfur and/or invest in costly flue gas desulfurization systems. Therefore, in the interest of overall national economics and preserving our limited Eastern supply of low sulfur coal, it is important that economic means be found and quickly implemented to encourage the utility industry to make use of our vast resources of higher sulfur-content coal.

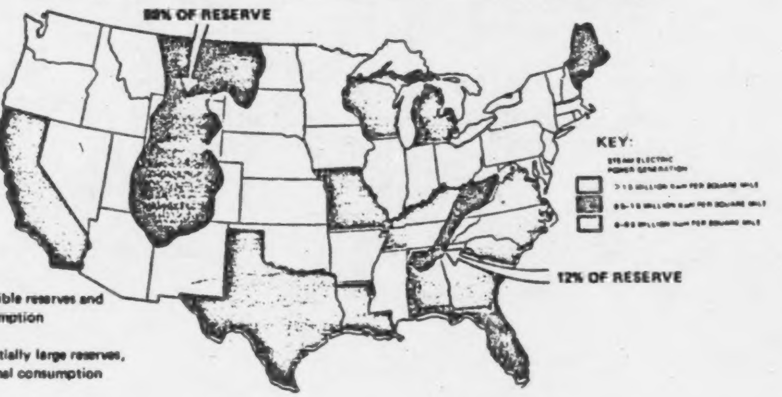
According to data published by the Bureau of Mines, the United States had minable underground and surface coal reserves of nearly 437 billion tons on 1 January 1974. Of this number, approximately 386 billion tons had been categorized according to sulfur content; 52% had 1% or less, 24% had between 1.0 and 3.0%, and 24% had greater than 3.0%. Although over half of these reserves fall into the low sulfur content range, 88% of the low sulfur coal is found west of the Mississippi (concentrated in such states as Montana and Wyoming), far removed from the electric power demand centers of the east. In the east only 24% of the coal is low sulfur. Therefore, in developing a reliable source of coal (long term supply not too far from point of consumption) to meet the demands of the eastern utilities, the medium and high sulfur content coals must be considered. It was the purpose of this study to identify an economic means of keeping such higher sulfur content coals in the energy market.

Data generated by the Bureau of Mines under the Federal Interagency Energy/Environment Research and Development Program indicate that some Eastern coals will show a significant reduction in ash and sulfur contents when physically cleaned to a 90% weight yield. With current technology, this level of physical cleaning can often be accomplished at an attractive cost. These coals with reduced ash and sulfur levels are often not too far removed from the sulfur content required to meet environmental standards in the geographic areas historically served by these coals.

When coal can be physically cleaned to a sulfur content close to that required to meet governing emission standards, a flue gas desulfurization (FGD) system treating only a portion of the flue gas would permit the coal burning facility to comply with such emission regulations. In many cases, the net cost of physical cleaning followed by FGD is substantially less than that associated with meeting standards exclusively with a larger capacity FGD system. This results from the fact that the savings associated with being able to use a smaller FGD system more than cover the net cost of physically cleaning the coal. Our study quantified this relationship under assumed conditions based upon current environmental standards being met by utility power plants using specific coals which are economically obtainable in the particular geographic areas considered.

The overall study findings covering new utility plants using physically cleaned coal followed by FGD indicated a savings of 2% to 112% as compared to meeting standards by FGD alone. The results were even more impressive for existing plants, where study assessments indicated a 13% to 140% savings for physical cleaning followed by FGD as compared to FGD alone.

LOW SULFUR COAL RESERVES VS. STEAM ELECTRIC POWER GENERATION



Hawaii: Negligible reserves and consumption

Alaska: Potentially large reserves, minimal consumption

Background

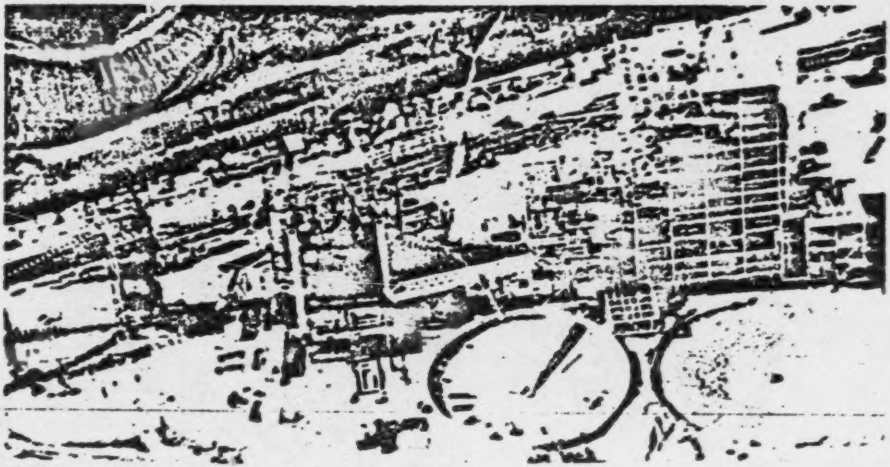
Coal is a heterogeneous material containing organic combustible matter and mineral matter. The mineral matter (i.e., impurities) may be broadly divided into two categories; those forming ash and those that contribute sulfur. Such ash-forming and sulfur-containing impurities may be further divided into two groups; 1) those that are chemically a part of the coal and cannot be removed by mechanical processes, and 2) those that are not chemically bound to the coal and can be removed to varying degrees by mechanical means. It is toward this latter category that physical coal cleaning (coal beneficiation, preparation, or washing) is directed.

Physical Cleaning

Physical cleaning is accomplished by first crushing the coal to liberate those impurities that are not chemically bound with the coal. These impurities are set free upon crushing since the seam where they are stuck to the coal is generally weaker than either the coal or the impurity. After the raw material (as mined coal) is crushed, the impurities can be separated out by mechanical means. The most common processes for achieving this separation are based upon the difference between the specific gravities of the impurities and that of the coal. Simply stated, coal has a specific gravity of around 1.3 which is less than that of its impurities. Thus, when the crushed particles (coal and impurities) are distended in water or other media a separation occurs as the heavier undesirable particles settle at a faster rate than the coal. In commercial scale cleaning operations, it is not uncommon to process 500 to 1000 tons of coal per hour using a variety of equipment designed and arranged around the makeup of the particular raw coal and the desired end product.

Physical cleaning of coal has been used for many years. In the past its principal purpose was to reduce the ash-forming impurities. However, today cleaning is of significant value in reducing the sulfur content of certain coals. Its applicability in this regard is not universal due to the various forms that sulfur occurs in coal. Sulfur exists in coal in two principal forms; organic and inorganic. Organic sulfur is one of those impurities referred to earlier which is chemically bound to the coal and cannot be removed by physical means. On the other hand, inorganic sulfur (pyritic sulfur) is not bound chemically and may be physically removed to varying degrees from the coal. The extent to which pyritic sulfur can be removed economically is a function of pyrite size and distribution. Once this information is obtained through careful laboratory analysis, then economic considerations will influence the level to which the coal is crushed and subjected to the cleaning operation.

Physical cleaning processes, while removing impurities from the coal, also reduce the total Btu available (i.e., a portion of the heat content of the raw coal is lost with



GPU's Homer City, Pa. Coal Cleaning Plant under construction the impurities). However the Btu content per unit weight of the cleaned coal increases due to the removal of the lower heat value impurities. In practice, an economic balance must be achieved between the Btu loss and the improvement in coal quality for various coals and applications. Certainly, this balance is further influenced by market and environmental considerations.

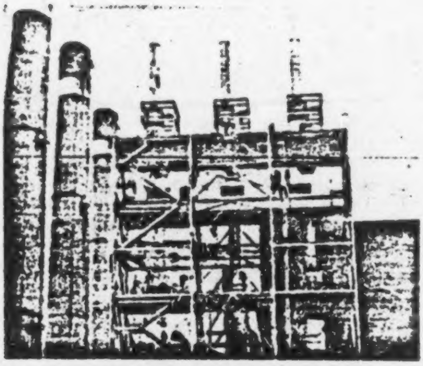
Problems

With regard to the environmental aspects, the oxides of sulfur (SO_x) created when coals are burned have long been recognized as a real threat to both the ecosystem and human health. Federal Standards, designed to protect against serious harm, have been established for new electric power plants which burn coal. To meet these standards, several methods are available to control sulfur oxides emissions from coal-fired combustion sources. These control methods include:

1. The use of low sulfur coal, either naturally occurring or physically cleaned;
2. Chemical treatment to extract sulfur from coal;
3. Removal of sulfur oxides from the combustion flue gas (stack gas scrubbing);
4. Conversion of coal to clean fuel by such processes as gasification and liquefaction.

Of these methods, for certain coals, physical cleaning to reduce sulfur content is the lowest cost and has the most developed technology. However, as noted earlier, physical cleaning can remove significant amounts of inorganic sulfur (pyritic) from certain coals, but has no influence upon the organic sulfur content. Therefore, when trying to come up with an environmentally acceptable product, physical cleaning may only be a partial step. This leads to another approach to controlling SO_x emissions, which could be the combined use of physical cleaning followed by stack gas scrubbing (i.e., flue gas desulfurization).

The purpose of a flue gas desulfurization (FGD) system is to limit the sulfur compounds escaping into the atmosphere. One of the most common FGD systems accomplishes its mission by forcing the SO_x in the flue gas to react with a limestone slurry. The cost of these systems goes up substantially with the amount of flue gas treated, which, in turn, is a function of the sulfur concentration of the flue gas stream. Therefore, significant savings can be realized if the sulfur concentration of the stream can be limited to the point where less than half of the total flue gas needs to be treated by the FGD system to meet standards. This becomes the basic economic concept of combining the two sulfur reduction technologies since such a combined approach could mean the ability to use higher sulfur content coals and minimal to moderate stack gas scrubbing.



Lime-Limestone Scrubber at TVA Power Plant

Physical Cleaning with Scrubbing

This concept of physical coal cleaning combined with flue gas desulfurization is not new. For some time there have been discussions, speculations, and some very preliminary assessments addressing the possible benefits of physical coal desulfurization followed by FGD. Past opinions based on a general appreciation of some of the cost and benefit factors associated with such an approach have led to the expressed belief that economic advantage in many instances could be attained. However, the associated specific economics had not previously been fully addressed. Therefore, to more completely define the potential economics associated with such a combined approach as a means of increasing the usefulness of some of our higher sulfur content coals, this study was initiated. It was conducted by the Bureau of Mines under the auspices of the EPA-coordinated Federal Interagency Energy/Environment R&D Program.

For purposes of this study, higher sulfur coals from the Northern Appalachian and Eastern Interior Regions were selected since they have been shown to have reasonable physical cleaning potential (i.e., economic ash and pyritic sulfur reduction). Then, possible users of these coals in the electric power generating industry were identified along with the current environmental constraints in their respective localities. This realism provided the framework within which to study and compare the economics associated with meeting sulfur emission standards in two alternative ways. The study considered both new and existing plants using either physically cleaned coal followed by stack gas scrubbing, or sulfur clean-up exclusively by stack gas scrubbing.

Economic Analyses

In order to determine the relative economics of meeting environmental standards via two different approaches, the factors influencing the costs and benefits of each element were identified. These are summarized as Table I.

For physical coal cleaning, the costs are principally dependent upon such factors as coal composition and plant size/cost. Offsetting these costs somewhat are the benefits of using cleaned coal. The magnitude of these benefits is influenced most by the increased heat content of the processed product brought on by the removal of some of the low heat content impurities. In the case of flue gas desulfurization (FGD), costs are sensitive to the sulfur composition of the flue gas stream.

TABLE I
KEY COST/BENEFIT FACTORS

COST OF CLEANING COAL

–Amortization of Cleaning Plant Capital Cost
–Operation and Maintenance Cost of Cleaning Plant
–Cost of Raw Coal Lost in Cleaning Process
–Taxes and Insurance Costs of Cleaning Plant

COSTS OF FLUE GAS DESULFURIZATION

–Amortization of Flue Gas Desulfurization (FGD) System Capital Cost
–Fuel and Electricity Cost Associated with FGD System
–Operating and Maintenance Costs of FGD System, Including:

Limestone • Fixation Chemicals • Operating Labor
Maintenance (Labor and Materials) • Supplies • Overhead

BENEFITS OF USING CLEANED COAL

–Increased Heat Content of Cleaned Coal
(Greater BTU Content Per Unit Weight)
–Transportation Savings (Less Weight to Ship for Same BTU Content)
–Ash Disposal Cost Saving (Clean Coal Leaves Less Ash)
–Pulverizing Cost Savings
(Less Cleaned Coal Needs to be Pulverized for Same BTU Content Required to Meet Output)
–Boiler and Related Equipment Maintenance Savings
(Clean Coal Is Less Corrosive)

Having identified the key cost/benefit factors associated with the two approaches, case analyses were performed based, in part, upon the major study factors appearing in Table II. As appropriate, these factors were based upon actual industry performance and practice.

TABLE II
MAJOR STUDY FACTORS

- The coals considered were those for which the Bureau of Mines has performed float-sink analyses. Bureau of Mines estimates of ash and sulfur levels versus top size and yield values were used.
- The top size coal considered was 3/8 inch.
- Economic assessments were based on a range of raw coal selling prices and a range of mining costs for a ton of coal.
- A coal cleaning plant cost of $18,000 per ton hour input capacity for plants of 500 tons per hour or greater.
- A coal cleaning plant financed by a 15-year equal payment self-liquidating loan.
- A coal cleaning plant use factor of 38.5 percent (i.e., plant operates 260 days per year at 13 hours per day).
- A coal cleaning plant property tax and insurance level equal to two percent of the initial investment.
- A power plant ash disposal cost of $4 per ton.
- A power plant coal pulverizing cost of $0.50 per ton.
- Annual stack gas scrubber capital charges that are based on the original investment and the remaining boiler life.
- Stack gas scrubber size/cost values based on recent EPA funded studies and a size/cost exponential relationship of 0.8.
- Stack gas scrubber operating labor cost that does not vary with plant size (i.e., there is a minimum practical labor level).
- General power plant locations that are consistent with actual conditions.
- Coal producing areas serving assumed user locations consistent with past patterns.
- Coal transportation parameters consistent with historical patterns and economical coal delivery (to insure conservative economic relationships).
- An assumed power plant boiler size of 500 Mw with an annual utilization of 7000 hours for a new plant and 5000 hours for a 10 year old plant.
- Stack gas scrubber systems of the minimum "practical" sizes necessary to meet standards.

A sample case study is presented as Table III. The key element in showing economic advantage in favor of the combined approach is in the significant difference in FGD costs. This significant difference results from only 21% of the flue gas being processed by the FGD system as opposed to 85% when cleaned coal is not used. A graphic representation of these two approaches is presented in the diagram opposite.

TABLE III
CASE STUDY

Problem:

Determine most cost-effective approach for new coal burning utility plant to meet emission standard of 1.2 lbs SO_2 per million Btu (MBTU)

Case Conditions:

Coal Use Area: Tonawanda (Buffalo), New York
Coal Source Area: Cambria County, Pennsylvania Coalbed: Lower Freeport
Raw Coal Characteristics: 11.4% Ash, 2.4% Sulfur
Cleaned Coal Characteristics: 6.7% Ash, 1.01% Sulfur

Costs of Alternate Approaches to Meeting Standard:

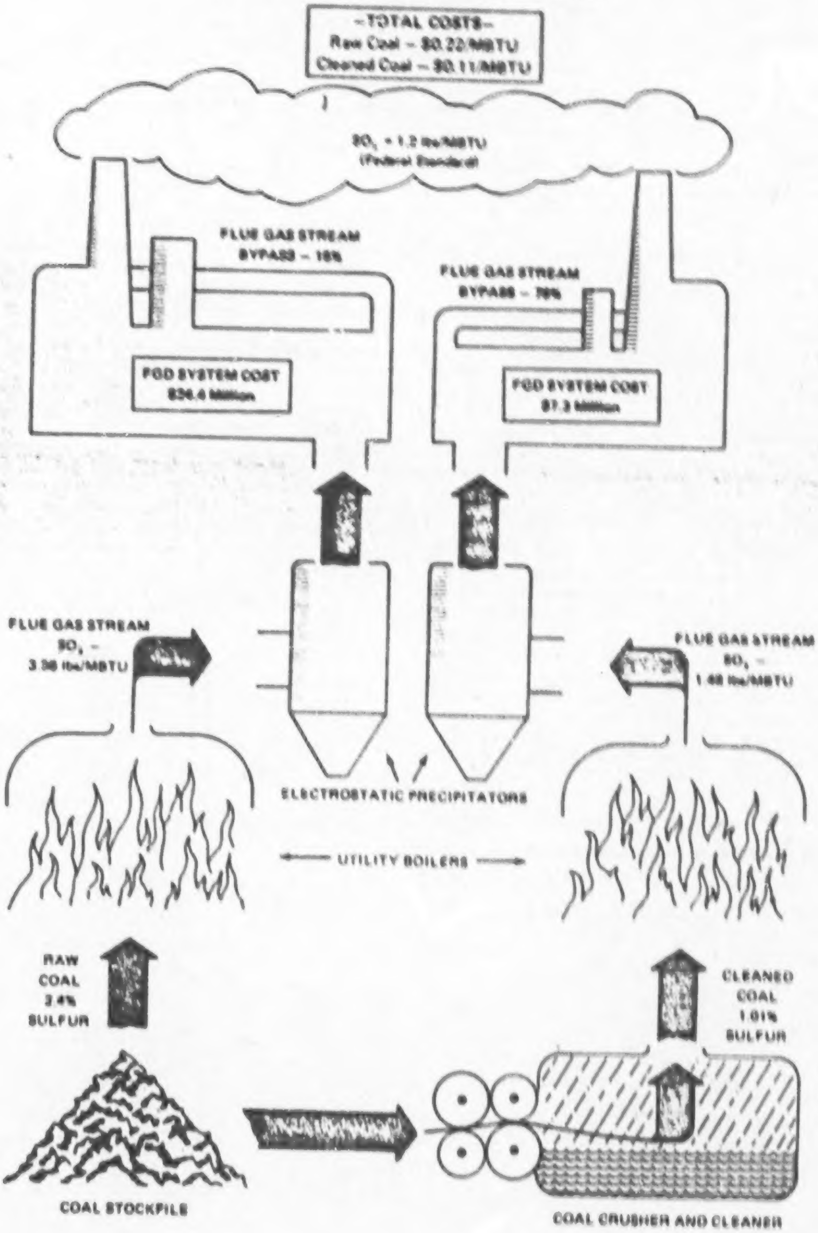
	Physical Cleaning Followed by FGD	*FGD Alone*
Coal Cleaning Cost		
Amortization of Cleaning Plant Capital Cost	$ 0.68/Ton	$ -0-
O & M Cost of Cleaning Plant	0.75/Ton	-0-
Cost of Coal Lost During Cleaning	1.56 To 2.00/Ton	-0-
Taxes & Insurance of Cleaning Plant	0.12/Ton	-0-
Total Cleaning Cost	$ 3.11 To 3.55/Ton	$ -0-
Cost of Flue Gas Desulfurization (FGD)		
Amortization of FGD Capital Cost	$ 0.92/Ton	$2.92/Ton
Fuel & Electricity of FGD System	0.17/Ton	0.66/Ton
O & M Cost of FGD System	0.72/Ton	2.24/Ton
Total FGD Cost	$ 1.81/Ton*	$5.82/Ton*
Benefits of Using Cleaned Coal		
Increased Heat Content	$ 1.21 To 1.40/Ton	$ -0-
Transportation Savings	0.30/Ton	-0-
Ash Disposal Savings	0.21/Ton	-0-
Pulverizing Savings	0.02/Ton	-0-
Maintenance & Other Savings	0.23/Ton	-0-
Total Benefit of Cleaning	$ 1.97 To 2.16/Ton	$ -0-
Net Cost (Costs Less Benefits)	$ 2.76 To 3.39/Ton	$5.82/Ton
Converted to per MBTU	$ 0.10 To 0.12	$0.22

SOLUTION:

The combined approach is the most cost-effective, compared to a 100% additional cost for meeting the standard by FGD alone.

*Significant difference in cost of FGD results from 21% of the flue gas being processed versus 85% in the more expensive approach to meet the required emission standard.

PHYSICAL COAL CLEANING WITH FLUE GAS DESULFURIZATION VS. FLUE GAS DESULFURIZATION ALONE


–TOTAL COSTS–
Raw Coal – $0.22/MBTU
Cleaned Coal – $0.11/MBTU
SO₂ = 1.2 lbs/MBTU
(Federal Standard)
FLUE GAS STREAM
BYPASS – 16%
FLUE GAS STREAM
BYPASS – 78%
FGD SYSTEM COST
$26.4 Million
FGD SYSTEM COST
$7.3 Million
FLUE GAS STREAM
SO₂ –
3.36 lbs/MBTU
FLUE GAS STREAM
SO₂ –
1.68 lbs/MBTU
ELECTROSTATIC PRECIPITATORS
UTILITY BOILERS
RAW
COAL
2.4%
SULFUR
CLEANED
COAL
1.01%
SULFUR
COAL STOCKPILE
COAL CRUSHER AND CLEANER

Additional Benefits

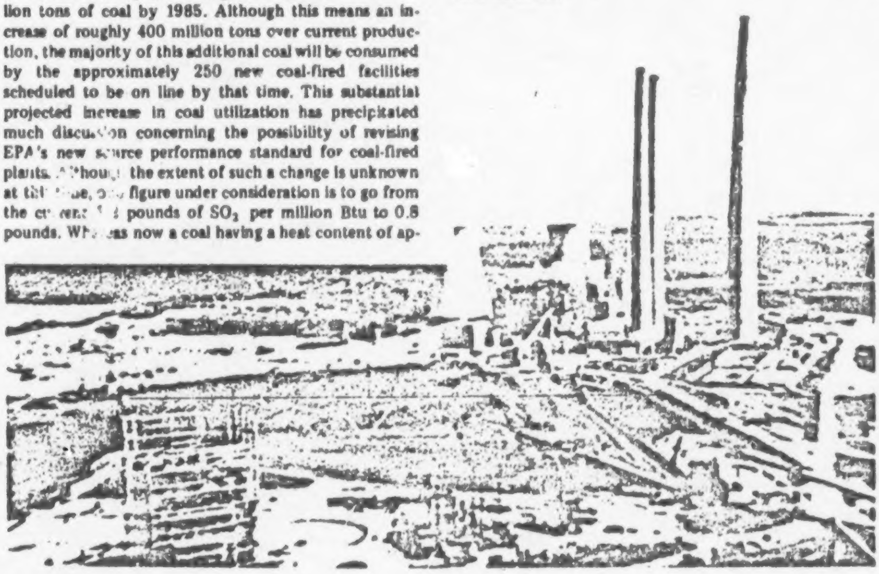
In addition to the benefits covered by this study, there are other positive features associated with combining physical desulfurization and FGD to meet emission standards. Such a benefit accruing from the combination of these two technologies is a decrease in the time required for FGD installation. This occurs for two reasons. First, the total sludge disposal at the plant site can be reduced to about one-quarter of what it would be in the absence of coal cleaning. This will, in some cases, reduce the time required for, and the difficulty with, obtaining legal permits and site acquisition for ponding and/or landfill. Second, as covered by this study, the cleaned coal reduces the volume of flue gas which must be scrubbed to meet emission standards, thereby requiring a smaller, or fewer, FGD units with an associated reduction in construction requirements and time.

Another benefit is the more intangible aspect of physical cleaning in the event of FGD shut-down. Although it cannot be readily quantified, there is a definite environmental benefit associated with the use of physically cleaned coal in the event of a failure in the SO_2 scrubbing system. This benefit is in the form of lower toxic emissions, particularly SO_2, than would have been the case if some sulfur and other contaminants had not been previously removed through physical cleaning.

Continued Prospects

The Federal government as well as the coal industry has established a production goal of more than one billion tons of coal by 1985. Although this means an increase of roughly 400 million tons over current production, the majority of this additional coal will be consumed by the approximately 250 new coal-fired facilities scheduled to be on line by that time. This substantial projected increase in coal utilization has precipitated much discussion concerning the possibility of revising EPA's new source performance standard for coal-fired plants. Although the extent of such a change is unknown at this time, one figure under consideration is to go from the current 1.2 pounds of SO_2 per million Btu to 0.8 pounds. Whereas now a coal having a heat content of approximately 25 million Btu's per ton and a sulfur content of 0.8% can meet standard, only those coals of 0.5% or less sulfur content could meet the 0.8 pounds of SO_2 per million Btu standard. If the standard is revised in this manner, there will be a significant reduction in the amount of coal which can be consumed without the use of SO_2 emission control devices. This would include both the coals naturally occurring with a low enough sulfur content and those capable of meeting the current standard through physical cleaning alone.

However, should the standard be amended, it will not negate the benefits associated with coal cleaning as set forth in this report (Table I). What it will mean is a reevaluation of emission control strategy on the part of potential coal users. Since coal having a sulfur content of 0.5% or less (either occurring naturally or after cleaning) is in limited supply and high in price due to pressure from the metallurgical market, it is impractical to consider the widespread burning of coal in new facilities without any SO_2 emission control devices under a 0.8 pounds of SO_2 per million Btu standard. This being the case, in searching for the most cost-effective approach to meeting any revised standard, the impact of coal preparation on reducing emission control costs as outlined by this report should be carefully considered. After such appraisal, users may find that, under their particular circumstances, the use of physically cleaned coal in combination with stack gas scrubbing will show a definite economic advantage even under more restrictive emission standards.



Summarization of Case Results

A total of forty-eight case analyses were performed. Each case examines one of twelve selected coals and possible use areas from the standpoint of both a new and an existing utility plant using either a combination of physically cleaned coal followed by stack gas scrubbing, or sulfur cleanup exclusively by stack gas scrubbing (i.e., flue gas desulfurization). The cases are grouped according to coal and use area. This permits ease of comparison between like plants using the same coal in the same area but utilizing two different approaches to meeting existing or projected environmental standards. In most cases, the emission standards used were applicable as of January 1976. However, where knowledge of projected changes was present, those standards were used and identified accordingly.

Included as part of this paper are summaries of each of the forty-eight cases analyzed. As stated above, these summaries are grouped in sets for comparison of the costs associated with meeting emission standards by the two alternate approaches. For example, Case Numbers 1A, 1B, 1C and 1D are based upon an actual coal coming from Sullivan County, Indiana which is assumed being used in a utility plant in the Knoxville, Tennessee area. Cases 1A and 1B approach the analysis on the basis of an assumed new facility, whereas Cases 1C and 1D assume an existing facility. However, Case Numbers 1A and 1C address the analysis using a combination of physically cleaned coal followed by stack gas scrubbing, whereas Case Numbers 1B and 1D approach the situation from the standpoint of using stack gas scrubbing alone. As can be readily seen from the summarization of these cases, the cost to meet the applicable sulfur emission standard is less in Case Numbers 1A and 1C which are the plans using physically cleaned coal followed by flue gas desulfurization (FGD) in new and existing facilities, respectively. In each set of cases, the relative economic advantage (or disadvantage) is expressed as a percent for comparative purposes. The manner of stating economic advantage (*less* costly) or disadvantage (*more* costly) is dependent upon what if any action the coal-using plant has taken to meet environmental standards. For example if, as in the case of 1C, the utility is using physically cleaned coal followed by FGD, then economic advantage could be stated as their cost is 25% less than by FGD alone. If, as in case 1D, the utility is using FGD alone, then economic disadvantage could be stated as their cost is 33% more than by the combined FGD/coal cleaning approach.

CASE RESULTS
[Sample Form]

Case Numbers:

Case Conditions

Coal Use Area: ____________________
Coal Source Area: ____________________ Coalbed: ____________
Raw Coal Characteristics: ______ % Ash ______ % Sulfur
Clean Coal Characteristics: ______ % Ash ______ % Sulfur

Comparison of Costs

CASE NO.	*TYPE OF PLANT & APPROACH*	*EMISSION STANDARD*	*COST TO MEET EMISSION STD.*	*ECONOMIC ADVANTAGE*
_	New Plant PC Followed by FGD			
_	New Plant FGD Alone			
_	Existing Plant PC Followed by FGD			
_	Existing Plant FGD Alone			

CASE RESULTS

Case Numbers: 1A, 1B, 1C, & 1D

Case Conditions

Coal Use Area: Knoxville (Clinton), Tennessee
Coal Source Area: Sullivan County, Indiana Coalbed: Number VII
Raw Coal Characteristics: 10.5 % Ash 1.87 % Sulfur
Clean Coal Characteristics: 7.3 % Ash 1.11 % Sulfur

Comparison of Costs

CASE NO.	*TYPE OF PLANT & APPROACH*	*EMISSION STANDARD*	*COST TO MEET EMISSION STD.*	*ECONOMIC ADVANTAGE*
1A –	New Plant PC Followed by FGD	1.2 lbs SO_2 per MBTU	\$0.15-0.17 per MBTU	16% less than by FGD alone
1B –	New Plant FGD Alone	1.2 lbs SO_2 per MBTU	\$0.19 per MBTU	19% more than by PC & FGD
1C –	Existing Plant PC Followed by FGD	1.2 lbs SO_2 per MBTU	\$0.23-0.25 per MBTU	25% less than FGD alone
1D –	Existing Plant FGD Alone	1.2 lbs SO_2 per MBTU	\$0.32 per MBTU	33% more than by PC & FGD

Case Numbers: 2A, 2B, 2C, & 2D

Case Conditions

Coal Use Area: Tonawanda (Buffalo), New York
Coal Source Area: Cambria County, Pennsylvania Coalbed: Lower Freeport
Raw Coal Characteristics: 11.4 % Ash 2.4 % Sulfur
Clean Coal Characteristics: 6.7 % Ash 1.01 % Sulfur

Comparison of Costs

CASE NO.	*TYPE OF PLANT & APPROACH*	*EMISSION STANDARD*	*COST TO MEET EMISSION STD.*	*ECONOMIC ADVANTAGE*
2A –	New Plant PC Followed by FGD	1.2 lbs SO_2 per MBTU	\$0.10-0.12 per MBTU	50% less than by FGD alone
2B –	New Plant FGD Alone	1.2 lbs SO_2 per MBTU	\$0.22 per MBTU	100% more than by PC & FGD
2C –	Existing Plant PC Followed by FGD	1.4 lbs SO_2 per MBTU	\$0.06-0.09 per MBTU	50% less than by FGD alone
2D –	Existing Plant FGD Alone	1.4 lbs SO_2 per MBTU	\$0.15 per MBTU	100% more than by PC & FGD

CASE RESULTS

Case Numbers: 3A, 3B, 3C & 3D

Case Conditions

Coal Use Area: Essexville (Saginaw), Michigan
Coal Source Area. Harrison County, Ohio Coalbed: Lower Freeport
Raw Coal Characteristics: 10.4 % Ash 2.30 % Sulfur
Clean Coal Characteristics: 4.8 % Ash 1.26 % Sulfur

Comparison of Costs

CASE NO.	*TYPE OF PLANT & APPROACH*	*EMISSION STANDARD*	*COST TO MEET EMISSION STD.*	*ECONOMIC ADVANTAGE*
3A –	New Plant PC Followed by FGD	1.2 lbs SO_2 per MBTU	\$0.12-0.14 per MBTU	38% less than by FGD alone
3B –	New Plant FGD Alone	1.2 lbs SO_2 per MBTU	\$0.21 per MBTU	62% more than by PC & FGD
3C –	Existing Plant PC Followed by FGD	1.6 lbs SO_2 per MBTU	\$0.11-0.14 per MBTU	58% less than FGD alone
3D –	Existing Plant FGD Alone	1.6 lbs SO_2 per MBTU	\$0.30 per MBTU	140% more than by PC & FGD

Case Numbers: 4A, 4B, 4C, & 4D

Case Conditions

Coal Use Area: Boston, Massachusetts
Coal Source Area: Clearfield County, Pennsylvania Coalbed: Upper Kittanning
Raw Coal Characteristics: 9.3 % Ash [illegible] % Sulfur
Clean Coal Characteristics: 7.0 % Ash 0.45 % Sulfur

Comparison of Costs

CASE NO.	*TYPE OF PLANT & APPROACH*	*EMISSION STANDARD*	*COST TO MEET EMISSION STD.*	*ECONOMIC ADVANTAGE*
4A –	New Plant PC Followed by FGD	0.28 lbs sul. per MBTU	\$0.12-0.14 per MBTU	24% less than by FGD alone
4B –	New Plant FGD Alone	0.28 lbs sul. per MBTU	\$0.17 per MBTU	31% more than by PC & FGD
4C –	Existing Plant PC Followed by FGD	0.28 lbs sul. per MBTU	\$0.17-0.19 per MBTU	38% less than by FGD alone
4D –	Existing Plant FGD Alone	0.28 lbs sul. per MBTU	\$0.29 per MBTU	61% more than by PC & FGD

CASE RESULTS

Case Numbers: 5A, 5B, 5C, & 5D

Case Conditions

Coal Use Area: Grand Rapids, Michigan
Coal Source Area: Preston County, W. Virginia Coalbed: Upper Freeport
Raw Coal Characteristics: 18.5 % Ash 2.24 % Sulfur
Clean Coal Characteristics: 11.9 % Ash 1.25 % Sulfur

Comparison of Costs

CASE NO.	*TYPE OF PLANT & APPROACH*	*EMISSION STANDARD*	*COST TO MEET EMISSION STD.*	*ECONOMIC ADVANTAGE*
5A –	New Plant PC Followed by FGD	1.6 lbs SO_2 per MBTU	$0.07-0.10 per MBTU	53% less than by FGD alone
5B –	New Plant FGD Alone	1.6 lbs SO_2 per MBTU	$0.18 per MBTU	112% more than by PC & FGD
5C –	Existing Plant PC Followed by FGD	1.6 lbs SO_2 per MBTU	$0.12-0.14 per MBTU	57% less than FGD alone
5D –	Existing Plant FGD Alone	1.6 lbs SO_2 per MBTU	$0.30 per MBTU	131% more than by PC & FGD

Case Numbers: 6A, 6B, 6C, & 6D

Case Conditions

Coal Use Area: Springfield, Massachusetts
Coal Source Area: Armstrong County, Penn. Coalbed: Upper Freeport
Raw Coal Characteristics: 13.0 % Ash 2.53 % Sulfur
Clean Coal Characteristics: 7.2 % Ash 1.09 % Sulfur

Comparison of Costs

CASE NO.	*TYPE OF PLANT & APPROACH*	*EMISSION STANDARD*	*COST TO MEET EMISSION STD.*	*ECONOMIC ADVANTAGE*
6A –	New Plant PC Followed by FGD	0.55 lbs sulfur per MBTU	$0.11-0.13 per MBTU	48% less than by FGD alone
6B –	New Plant FGD Alone	0.55 lbs sulfur per MBTU	$0.23 per MBTU	92% more than by PC & FGD
6C –	Existing Plant PC Followed by FGD	0.55 lbs sulfur per MBTU	$0.18-0.21 per MBTU	49% less than by FGD alone
6D –	Existing Plant FGD Alone	0.55 lbs sulfur per MBTU	$0.38 per MBTU	95% more than by PC & FGD

CASE RESULTS

Case Numbers: 7A, 7B, 7C, & 7D

Case Conditions

Coal Use Area: Lansing, Michigan

Coal Source Area: Jefferson County, Ohio Coalbed: Pittsburgh

Raw Coal Characteristics: 9.8 % Ash 2.82 % Sulfur

Clean Coal Characteristics: 6.0 % Ash 2.03 % Sulfur

Comparison of Costs

CASE NO.	*TYPE OF PLANT & APPROACH*	*EMISSION STANDARD*	*COST TO MEET EMISSION STD.*	*ECONOMIC ADVANTAGE*
7A –	New Plant PC Followed by FGD	1.6 lbs SO_2 per MBTU	$0.19-0.22 per MBTU	6.8% less than by FGD alone
7B –	New Plant FGD Alone	1.6 lbs SO_2 per MBTU	$0.22 per MBTU	7.3% more than by PC & FGD
7C –	Existing Plant PC Followed by FGD	1.6 lbs SO_2 per MBTU	$0.30-0.32 per MBTU	11% less then FGD alone
7D –	Existing Plant FGD Alone	1.6 lbs SO_2 per MBTU	$0.35 per MBTU	13% more than by PC & FGD

Case Numbers: 8A, 8B, 8C, & 8D

Case Conditions

Coal Use Area: Nashville (Gallatin), Tennessee

Coal Source Area: Vigo County, Indiana Coalbed: Number VII

Raw Coal Characteristics: 12.0 % Ash 1.54 % Sulfur

Clean Coal Characteristics: 7.7 % Ash 0.90 % Sulfur

Comparison of Costs

CASE NO.	*TYPE OF PLANT & APPROACH*	*EMISSION STANDARD*	*COST TO MEET EMISSION STD.*	*ECONOMIC ADVANTAGE*
8A –	New Plant PC Followed by FGD	1.2 lbs SO_2 per MBTU	$0.11-0.13 per MBTU	25% less than by FGD alone
8B –	New Plant FGD Alone	1.2 lbs SO_2 per MBTU	$0.16 per MBTU	33% more than by PC & FGD
8C –	Existing Plant PC Followed by FGD	1.2 lbs SO_2 per MBTU	$0.16-0.18 per MBTU	39% less than by FGD alone
8D –	Existing Plant FGD Alone	1.2 lbs SO_2 per MBTU	$0.28 per MBTU	65% more than by PC & FGD

CASE RESULTS

Case Numbers: 9A, 9B, 9C, & 9D
Case Conditions
Coal Use Area: Burlington, New Jersey
Coal Source Area: Garrett County, Maryland Coalbed: Upper Freeport
Raw Coal Characteristics: 13.8 % Ash 2.37 % Sulfur
Clean Coal Characteristics: 8.8 % Ash 1.6 % Sulfur

Comparison of Costs

CASE NO.	*TYPE OF PLANT & APPROACH*	*EMISSION STANDARD*	*COST TO MEET EMISSION STD.*	*ECONOMIC ADVANTAGE*
9A –	New Plant PC Followed by FGD	0.30 lbs SO_2 per MBTU	This coal will not meet emission standards for new plants in the State of New Jersey with either combined physical cleaning and FGD or FGD alone.	
9B –	New Plant FGD Alone	0.30 lbs SO_2 per MBTU		
9C –	Existing Plant PC Followed by FGD	1% sulfur by weight per MBTU	$0.23-0.25 per MBTU	25% less than FGD alone
9D –	Existing Plant FGD Alone	1% sulfur by weight per MBTU	$0.32 per MBTU	33% more than by PC & FGD

Case Numbers: 10A, 10B, 10C, & 10D
Case Conditions
Coal Use Area: Milwaukee, Wisconsin
Coal Source Area: Franklin County, Illinois Coalbed: Number 6
Raw Coal Characteristics: 14.8 % Ash 1.12 % Sulfur
Clean Coal Characteristics: 7.1 % Ash 0.95 % Sulfur

Comparison of Costs

CASE NO.	*TYPE OF PLANT & APPROACH*	*EMISSION STANDARD*	*COST TO MEET EMISSION STD.*	*ECONOMIC ADVANTAGE*
10A –	New Plant PC Followed by FGD	1.2 lbs SO_2 per MBTU	$0.07-0.10 per MBTU	29% less than by FGD alone
10B –	New Plant FGD Alone	1.2 lbs SO_2 per MBTU	$0.12 per MBTU	41% more than by PC & FGD
10C –	Existing Plant PC Followed by FGD	1.2 lbs SO_2 per MBTU	$0.12-0.15 per MBTU	33% less than by FGD alone
10D –	Existing Plant FGD Alone	1.2 lbs SO_2 per MBTU	$0.20 per MBTU	48% more than by PC & FGD

CASE RESULTS

Case Numbers: 11A, 11B, 11C, & 11D

Case Conditions

Coal Use Area: Concord, New Hampshire

Coal Source Area: Greene County, Pennsylvania Coalbed: Sewickley

Raw Coal Characteristics: 11.4 % Ash 3.45 % Sulfur

Clean Coal Characteristics: 8.1 % Ash 2.20 % Sulfur

Comparison of Costs

CASE NO.	*TYPE OF PLANT & APPROACH*	*EMISSION STANDARD*	*COST TO MEET EMISSION STD.*	*ECONOMIC ADVANTAGE*
11A –	New Plant PC Followed by FGD	1.2 lbs SO_2 per MBTU (projected standard)	\$0.25-0.28 per MBTU	2% less than by FGD alone
11B –	New Plant FGD Alone	1.2 lbs SO_2 per MBTU (projected standard)	\$0.27 per MBTU	2% more than by PC & FGD
11C –	Existing Plant PC Followed by FGD	1.5 lbs sulfur per MBTU	\$0.12-0.14 per MBTU	54% less than FGD alone
11D –	Existing Plant FGD Alone	1.5 lbs sulfur per MBTU	\$0.28 per MBTU	115% more than by PC & FGD

Case Numbers: 12A, 12B, 12C, & 12D

Case Conditions

Coal Use Area: Dickerson (Montgomery County), Maryland

Coal Source Area: Marion County, W. Virginia Coalbed: Pittsburgh

Raw Coal Characteristics: 11.0 % Ash 3.80 % Sulfur

Clean Coal Characteristics: 5.9 % Ash 2.16 % Sulfur

Comparison of Costs

CASE NO.	*TYPE OF PLANT & APPROACH*	*EMISSION STANDARD*	*COST TO MEET EMISSION STD.*	*ECONOMIC ADVANTAGE*
12A –	New Plant PC Followed by FGD	1% sulfur by weight	\$0.20-0.23 per MBTU	20% less than by FGD alone
12B –	New Plant FGD Alone	1% sulfur by weight	\$0.27 per MBTU	26% more than by PC & FGD
12C –	Existing Plant PC Followed by FGD	1% sulfur by weight	\$0.32-0.34 per MBTU	23% less than by FGD alone
12D –	Existing Plant FGD Alone	1% sulfur by weight	\$0.43 per MBTU	30% more than by PC & FGD

FOR FURTHER READING

Engineering/Economic Analyses of Coal Preparation with SO_2 Clean Up Process for Keeping Higher Sulfur Coals In the Energy Market. Final report under U.S. Bureau of Mines, Contract no. J-0155-171. November 1976.

This report is a comprehensive study of 48 actual cases where either a combination of physical coal cleaning and scrubbing was used, or sulfur clean up was accomplished entirely by flue gas scrubbing. The full report will be published by EPA in mid-1977 and will be available from the Office of Energy, Minerals, and Industry, RD-681, U.S. EPA, Washington, D.C. 20460.

Sulfur Reduction Potential of the U.S. Coals: A Revised Report of Investigations EPA-600/2-76-091

This report presents the results of a washability study of 455 raw coal samples from major United States coal production regions. The emphasis was on sulfur reduction potential with summary data of the number of samples meeting the EPA new source performance standard under varying degrees of preparation (cleaning).

Steam—Electric Plant Factors, 1975 Edition, National Coal Association, Washington, D.C. 20036.

From this NCA publication, existing and planned electric power plant generation capacity can be observed by specific location. Additionally the type of fuel being consumed by the specific plant is given with heat content and price. From this information, future demand patterns can be projected.

Demonstrated Coal Reserve Base of the United States, By Sulfur Category, On January 1, 1974. U.S. Bureau of Mines, Mineral Industry Surveys, Washington, D.C. 20241. May 1975.

This report presents a concise geographic summary of U.S. coal reserves according to three ranges of sulfur content.

"All That An Oil Company President Needs To Know About Washing Bituminous Coal," Special Issue of *B.T.U.*, McNally Pittsburg Manufacturing Corp., Pittsburg, Kansas, 66762. May 1972.

An extremely well-written, non-technical description of the major coal cleaning methods. Also includes easy to understand definitions of industry terms.

The Effect of Coal Quality on the Operation and Maintenance of Large Central Station Boilers. Paper written by John G. Holmes, Jr. of TVA for presentation at Annual Meeting of American Institute of Mining, Metallurgical, and Petroleum Engineers, Washington, D.C. February 16-20, 1969.

Presents comparative data on maintenance records of two boilers utilizing coals of different ash and sulfur contents. These data demonstrate a definite reduction in O & M costs as a function of reduced impurities. This publication may be obtained by writing to American Institute of Mining, Metallurgical and Petroleum Engineers, Inc., 345 E. 47th Street, N.Y., N.Y. 10017.

This article from Power magazine reviews processes that physically clean coal; listing the benefits and problems associated with coal cleaning. Coal cleaning, in combination with Flue Gas Desulfurization, can reduce the overall cost of SO2 removal. Table 3 (p. 126 presents the various washing processes and their efficiencies.

Power
September 1981
Volume 125
Number 9

A special report

Coal cleaning: improve boiler performance and reduce SO_2 emissions

By Paul S Baur, Assistant Editor

If you are an engineer with experience in fossil-fueled steam and power production for a utility or a process plant, chances are that the term coal preparation means something different to you than it would to a mining engineer involved in coal production. The power specialist is most familiar with that segment of preparation performed immediately prior to combustion—the crushing, drying, and pulverizing required to provide uniform fuel flow to the boiler.

Coal preparation at a mine has a somewhat broader definition; it includes anything from simple crushing and sizing to extensive beneficiating processes used to improve the quality of the coal without destroying its physical and chemical identity. Physical coal cleaning (a term synonymous with physical beneficiation, physical coal preparation, and coal washing), as it is commonly practiced today, is a water-based process that concentrates combustible materials by removing extraneous ash and other impurities through specific-gravity separation. The main objective of most steam-coal cleaning processes is to strike an economic balance between high carbon recovery and reduced ash and sulfur levels.

Physical cleaning is different from chemical coal cleaning, which refers to processes specifically designed to remove primarily sulfur by chemical reaction. Although most chemical cleaning processes can remove a higher percentage of sulfur than physical cleaning, they are not considered economically competitive with current methods of reducing sulfur emissions such as washing, direct combustion of low-sulfur coal, and flue-gas desulfurization.

As a current or prospective coal user, you should be aware of the benefits preparation technology offers, as well as some of its limitations. At present, only 15-20% of the nearly 600-million tons of coal consumed yearly by the utility industry is cleaned, and generally only to a small degree; the rest is burned run-of-mine (r-o-m). Developing circumstances, however, are making coal cleaning more economical. They are:

- Higher coal prices and transportation costs.
- Diminishing coal quality, due to less-selective mining techniques and more difficult-to-mine seams.
- Utilities' desire to increase generating capacity without the need for major capital investment.

■ Increasingly stringent air-quality standards.

Historically, there have been few incentives for using cleaned coal, because air-pollution standards did not become restrictive until the 1970s, and the economic benefits of reduced ash and sulfur on boiler performance were difficult to quantify. Even today, plants are just beginning to document the savings in operation and maintenance costs from using clean coal.

Other reasons for cleaning's relatively slow acceptance are based on traditional utility attitudes (box, page S-3). Most companies have subscribed to the philosophy that the cheapest fuel produces the cheapest energy. However, many of the large coal users—such as American Electric Power and the Tennessee Valley Authority—are finding that this is not always true. Mounting evidence shows that cleaned coal produces steam and electricity at lower cost than r-o-m coal.

Benefits clarified

Now let's look at some of the benefits in detail:

■ **Reduced transportation costs.** Over half of all steam coal produced in the US is shipped to the user by rail. Since beneficiation can eliminate waste that accounts for 15-35% of the mass of most r-o-m coal, a significant long-term saving can be realized where hauls are long and the volumes handled are large. Be aware, however, that the higher surface moisture of washed coal may partially negate the benefits of ash reduction, and may also aggravate a freeze-control problem during the winter.

Provisions of most United Mine Worker contracts require payment by the mine operator to the UMW pension and trust fund for every ton of coal shipped. Thus, for utilities with captive mines, the reduced shipping tonnage due to washing could represent a significant saving. One other point: The slurry-pipeline coal-transport concept and coal cleaning fit together like hand-in-glove; slurrying will make cleaning at the mine more practical, if not an economical necessity.

■ **Improved coal quality.** Washing eliminates some of coal's noncombustible mineral matter, and also reduces the fuel to a uniform size. This creates a higher-quality, higher-Btu product, which costs less to transport, handle, crush, and pulverize at the plant. In additon, the higher Btu value lowers the required fuel-burn rate for firing equipment, increases the ability of a boiler to meet peak-load demand, and eliminates fluctuations in plant heat rate resulting from the variability normally found in r-o-m coal.

Cleaning also provides a method of combatting the gradual decline in the quality of r-o-m coal experienced over the past decade. Many of the nation's prime coal seams are being depleted through aggressive mining and the use of mechanized and continuous mining methods. These mining techniques are generally less selective, so they produce, along with the coal, a significant percentage of roof and floor material from the seam bed. Washed coal can save a plant from untimely outages caused by excessive wear on crushers and pulverizers due to these impurities.

■ **Improved boiler performance.** Physical beneficiation significantly alters ash quantity and the consistency of the coal, thereby reducing slagging, fouling, and corrosion-related problems. Many studies have attempted to determine to what extent the cost of upgrading fuel quality through beneficiation is offset by reductions in boiler operating and maintenance costs, and increased availability. Because of lack of historical data for utilities burning cleaned coal, however, the effects of ash and sulfur on power-plant operation have yet to be precisely quantified. It is generally agreed among leaders in this research, however, that even a small improvement in availability would justify the cost of cleaning coal in almost all cases.

■ **Less ash to handle.** Cleaning can reportedly remove up to 70% of the ash in some coals, reducing ash-handling and ash-disposal costs at the plant. In addition, the flyash loading on electrostatic precipitators (ESPs) and fabric-filter collectors can be reduced. Cleaned coal may even be a requirement at some plants where state or federal limits for particulates and flyash are stringent.

But be aware that ash reduction through cleaning is not always advantageous when using an ESP to control particulates. Reason: The lower sulfur level of cleaned coal increases flyash resistivity and hampers its capture. Though ESPs may have to be made larger in some cases where coal is cleaned, the overall decrease in flyash levels may be a partially compensating factor. Another factor: Cleaner coal means less sensible heat is lost to bottom ash in the boiler; the result is improved overall heat transfer.

■ **Trace-contaminant control.** The cleaning process partitions potentially harmful elements in the coal among

1. This utility-owned preparation plant in Homer City, Pa, yields two grades of coal, to meet separate emission limits at a minemouth plant located nearby

A final obstacle—overcoming investor reluctance

Why hasn't coal cleaning made greater inroads in utility and industry planning? The answer lies with potential investors—coal companies, utilities, and large industrial coal users. First of all, few investors are aware of the wide range of benefits to be reaped from clean coal. Also, they are unsure about the ultimate costs of pollution-control requirements as they apply both to the cleaning facility and to the powerplant, especially SO_2 emission standards. Small investors are not even certain they can raise the necessary capital without sharing the cost burden. Finally, the Internal Revenue Service still hasn't recognized precombustion sulfur removal as a candidate for investment credits, despite provisions in the Clean Air Act that permit them.

This atmosphere of uncertainty makes coal cleaning a risky investment in the minds of most potential investors. To overcome the barriers, more information about coal cleaning (especially cost savings) needs to be developed and disseminated. Recent efforts by government agencies and utilities are helping to spread the word, but an information gap still remains. Also, the area of joint ventures should be more fully explored to alleviate the impact of first cost. With private investments, this is usually straightforward; with utilities, however, it is more complex because of the stringent requirements of state public utility commissions.

Continued pressure should be applied to the IRS to reconsider its ruling. With coal cleaning qualifying as a tax credit, investors could accelerate equipment depreciation for tax purposes, claim a 5% investment tax credit, issue tax-exempt bonds to finance the plant, and apply for exemptions from state sales and use taxes.

With more information about coal cleaning, better financing arrangements, and a more attractive tax structure, the incentive to invest in coal cleaning would be much enchanced.

size/gravity fractions, and may be able to serve as a precombustion removal method. This could become more important as EPA performs studies on the harmful effects of such contaminants, and decides to control their release.

- **FGD operating costs.** Cleaning, when combined with flue-gas desulfurization as a method of SO_2 control, significantly reduces the volume of limestone consumed and of sludge produced by the FGD system. These scrubbers are particularly sensitive to the variability of sulfur and the presence of other contaminants found in raw coal; cleaning often alleviates this problem. Studies sponsored by EPA and other units have shown that physically cleaned coal lowers the total cost of SO_2 control.
- **Lower cost of new plants.** In a recently published study, the Electric Power Research Institute (EPRI) attempted to quantify the cost effects of using clean vs r-o-m coal to fuel new powerplants. Cost benefits were estimated for individual components of the power-generation system—such as transportation, plant operation, pollution control, and waste disposal. In most cases, the estimated savings in capital and operating costs more than offset the added cost of cleaning.

It's important to realize that any justification for using washed coal in a plant should be based on a host of factors that are specific to the particular plant site; no sweeping generalizations can be made. Before the decision is made to use cleaned coal, a detailed economic study should be able to quantify the expected costs and benefits of using that higher-quality fuel.

Cleaning to remove sulfur

When the 1970 amendments to the Clean Air Act mandated a decrease in SO_2 emissions, coal beneficiation to remove sulfur gained considerable momentum. Coal traditionally has been cleaned to remove ash, particularly in the steel industry, where a high-quality, uniform coal is required for coking. But because part of the sulfur in r-o-m coal is contained in its mineral matter, coal washing can remove this environmental contaminant as well.

Physical beneficiation cannot remove all of the sulfur associated with coal, because a significant amount of the sulfur usually is chemically bound up in the coal structure. But much of the unbound (so-called pyritic) sulfur can be removed by simple washing. In fact, as much as 50% of the coal's total sulfur content can be eliminated by physical beneficiation. Higher removal efficiencies, however, require chemical cleaning to remove the inherent organic sulfur.

Coal cleaning is considered the least expensive and most reliable method of removing pyritic sulfur. Washing reportedly reduces SO_2 emissions at a cost of $100-$150/ton of sulfur removed, a procedure that reportedly costs a fraction of that of conventional FGD.

Washing to meet SO_2 limits. As noted earlier, the three commercially viable methods of complying with existing or new-source SO_2 regulations are cleaning, FGD, or firing with low-sulfur coal. Compliance can be accomplished by using any one of these, or a combination of two or all three. In Illinios, for example, where high-sulfur coal is abundant, a combination of all three methods—using a blend of washed, locally mined coal with low-sulfur Western coal, plus FGD—may be necessary under some circumstances.

How your plant will best be able to meet state or federal emission guidelines depends on many factors, each specific to a particular location, coal source, and boiler type. Despite this ambiguity, however, some general conclusions can be made which may apply to your plant.

FGD combined with cleaning. You can expect that coal beneficiation will gain importance as a pollution-control method for *new* sources, when used in conjunction with FGD to meet the strict SO_2-reduction requirement defined by EPA. Though the 1977 revisions to the Clean Air Act require FGD on all new central stations, regardless of the sulfur content of the coal, there is strong reason to believe that the demand for cleaned coal will continue to increase. Reason: When clean coal is burned, the FGD system can be designed to process only a portion of the flue-gas stream. The remainder is bypassed, alleviating the need for plume reheat (often required to restore effluent buoyancy) and its associated energy costs.

In some cases, the net cost of physical cleaning followed by FGD is substantially less than the cost of larger-capacity FGD system. Some experts believe it may not be technically feasible to comply with present and future NSPS regulations solely by scrubbing, so that the use of other control technologies in tandem will be necessary.

Washing for credits. Under the revised NSPS for utility boilers built after Sept 18, 1978, any cleaning of the fuel that reduces its pollution characteristics after its extraction from the ground and prior to combustion may be credited to the pollution percentage-reduction requirement. In some cases, this credit would allow a utility that has purchased washed coal to reduce its scrubbing requirement (box, page S-4).

Some industry officials are skeptical, however, about the merits of combined cleaning and FGD. They argue that very few coals can have sufficient sulfur removed by cleaning to affect FGD designs significantly. To illustrate this point: If 30% of the sulfur is removed by cleaning, FGD must still remove nearly 80% of the remaining sulfur to meet an original 85% reduction requirement.

Cleaning could, however, improve scrubber performance considerably by reducing its removal-efficiency requirement. Experience shows that many FGD systems have been plagued by major operating problems, which could be alleviated by burning lower-sulfur coal.

Plants that are governed by less-stringent state emission regulations, as well as those under the old federal NSPS regulations—which did not include a mandatory percentage-reduction requirement, and had a ceiling of 1.2 lb $SO_2/10^6$ Btu for all boilers rated more than 250-million Btu/hr heat input—may be able to meet their emission requirement solely by firing washed coal. And more plants may fall into this category if emission regulations are relaxed by Congress in the next revision of the Clean Air Act, slated for later this year.

This is good news for many coal-mine

Cleaned coal receives credit in EPA sulfur-reduction requirement

The most recent revision of the New Source Performance Standards (promulgated in June 1979) requires all new utility boilers (construction beginning after Sept 18, 1978) with heat-input rates greater than 250 x 10^6 Btu/hr to limit SO_2 emissions as a function of the sulfur content of the coal being fired. These standards stipulate maximum emission limits for coal of 1.2 lb/10^6 Btu heat input to the boiler, plus a sliding-scale sulfur-reduction requirement of 90% for high-sulfur coal, down to a minimum 70% for low-sulfur coal (coal producing less than 0.6 lbs $SO_2/10^6$ Btu heat input during uncontrolled combustion). The regulations represent minimum acceptable emission levels for large sources.

Since their inception, the revised NSPS have been a constant source of misunderstanding. At first glance, they appear to result in a step function, with only two SO_2-control levels—one for high- and one for low-sulfur coal. In fact, however, a variable scale of SO_2-control requirements results (top figure).

For coals in region A of this figure (greater than 12 lb/10^6 Btu), the 1.20-lb/10^6 Btu emission ceiling is the controlling factor. In region B (6-12 lb/10^6 Btu), the 90% reduction requirement is controlling, and the resulting emissions will be less than 1.2 lb/10^6 Btu. For example, when firing a coal with a potential emission rate of 8 lb/10^6 Btu, a 90% reduction would result in an actual emission rate of 0.8 lb/10^6 Btu.

In region C (2-6 lb/10^6 Btu), the SO_2 requirements can be seen to vary continuously with the sulfur content of the coal. A plant burning coal with a potential emission rate of 6 lb/10^6 Btu must reduce emissions by 90% to meet an emission limit of 0.6 lb/10^6 Btu. On the other end of the scale, a plant burning a coal with a potential emission rate of 3 lb/10^6 Btu only needs an 80% SO_2 reduction to meet an emission rate of 0.6 lb/10^6 Btu.

Finally, for coals in region D, the minimum 70% reduction requirement is controlling, and the resulting emission is less than 0.6 lb/10^6 Btu.

The percentage-reduction requirement immediately rules out coal washing as a sole means of emission compliance. At best, cleaning can only remove 30-40% of the total sulfur content of coals that are high in sulfur to begin with; sulfur reduction through physical beneficiation of lower-sulfur coals is even less effective.

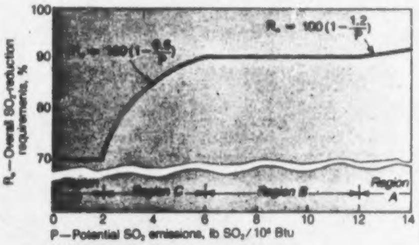
SO_2 reduction can, however, be achieved by any combination of precombustion and postcombustion methods, including washing, FGD, fluidized-bed combustion, etc. But don't make the mistake of attempting to add two percentage reductions; the overall SO_2 reduction must be calculated.

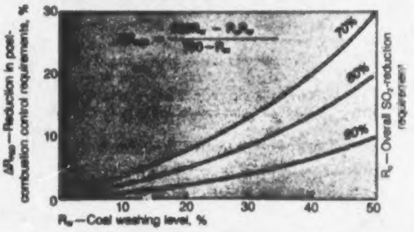
To see how this works, suppose a coal with a potential emission rate of 10 lb/10^6 Btu is being fired. An overall 90%-reduction requirement applies, resulting in an allowable emission rate of 1.0 lb/10^6 Btu. If the coal is washed to reduce the potential emission rate by 20%, this reduces the FGD requirement from 90% to 87.5% (not to 70%). The actual rate on leaving the washing process is 8.0 lb/10^6 Btu. To reduce this rate further, to 1.0 lb/10^6 Btu, needs an 87.5% reduction by FGD.

The calculation procedure allows full credit for emission reduction from coal washing, but does not reduce the FGD requirements on a one-to-one basis. The 20% washing of a 10-lb/10^6 Btu coal allows a 2.5% reduction in FGD control; 40% washing allows a 6.7% reduction in FGD. Washing credits are also a function of the overall SO_2 reduction requirements for the coal (bottom figure).

Credit for coal washing used toward reducing post-combustion removal requirements need not be claimed, because it is automatically accounted for by the sliding scale. This reduces recordkeeping and reporting requirements. For example, if a coal with a potential SO_2 emission rate of 7 lb/10^6 Btu were washed 40%, to a rate of 4.2 lb/10^6 Btu, it would automatically drop from the 90% SO_2-reduction bracket to a 70% bracket—with no need to claim the credit.

Using the above procedure, a combination of coal washing, FGD, and other technologies can be considered to select the most cost-effective control option in each case.





owners in the Midwest and Appalachia, who are in danger of being forced out of the steam-coal market because of the high sulfur content of their product. Much of this coal is relatively easy to clean, because half or more of the sulfur content is pyritic and can be removed by vigorous washing.

One Illinois seam, for example, contains coal with 4% total sulfur, roughly 60% of which is pyritic. If washed coal from this seam contains 2.2% sulfur and has a heating value of 11,000 Btu/lb, it would be able to meet an emissions limit of 4.0 lb $SO_2/10^6$ Btu heat input. The washing reduction not only allows existing facilities to continue burning coal from present sources, but also provides the least-expensive means of compliance.

Coal washing may also mean good news for those in the process of converting boilers to coal. The New England Electric System, which recently reconverted its Brayton Point plant from oil back to coal, is allowed by EPA to burn cleaned coal from Pennsylvania, Virginia, and West Virginia, instead of installing government-stipulated best-available control technology—which almost always means scrubbers. The plant, which is the largest coal-fired station in the region, has demonstrated that it can operate well within EPA guidelines on the washed coal. The decision to take this route reduced the capital cost of conversion roughly to half of the expected total, and has provided a significant reduction in the originally expected operating cost of using FGD.

NSPS guidelines for industrial boilers have not been promulgated yet. If they are similar to utility regulations, only the largest industrial units will be able to afford compliance. If the standards are arranged according to boiler capacity, as they are expected to, smaller units may be able to comply by burning cleaned coal. Until other pollution-control technologies have matured, such as fluidized-bed combustion, cleaned coal may be the most viable interim alternative for small boilers to meet the requirement.

Another incentive to reduce the sulfur content of coal by washing is a banking and trading scheme promoted by EPA. Through this program, both existing and new plants can receive credits from their state air-pollution-control agencies for reducing emissions more than the legal requirement. A utility may use the banked credit itself at a later date, to accommodate capacity expansion, or may sell it to another utility to help meet its emissions requirements and avoid the cost of installing pollution-control equipment. This credit scheme offers utilities in some areas of the country a measure of flexibility in responding to emission standards.

Principles

To design a coal-cleaning process, a preparation engineer generally requires two pieces of information: (1) An accurate physical and chemical characterization of the coal to be cleaned—determined from a washability analysis; and (2) the cleaned-coal product requirements specified by the end user, and ultimately determined by the preparation-plant owner, whether it be a mine, utility, or combined ownership. From this information, engineers draw on their knowledge, experience, and philosophy to determine the most practical method of designing the process. Remember, however, that plant design is an art rather than a science; given the same design criteria, no two engineers are likely to agree on precisely the same plant design.

Most cleaning processes rely on the relative differences in size, specific gravity, and surface properties of raw coal to achieve separation of clean coal and refuse. Specific-gravity separation is the most widely used method. Impurities usually have a specific gravity greater than 2, while coal ranges from 1.30 to 1.75 (see Table 1). When placed in a medium whose specific gravity is between these two, impurities will tend to sink while coal will float.

Key separation methods

Three methods are used to separate coal by specific gravity:

- Water-only separation is the most common, and is used with easy-to-separate coals. Equipment includes coarse- and fine-coal jigs, concentrating tables, and hydrocyclones.
- Heavy-medium vessels and heavy-medium cyclones use a liquid of a specific gravity between that of clean coal and refuse to achieve efficient separation of hard-to-clean coals that contain large amounts of "near-gravity" material (coal and mineral particles near in density to the specific gravity of separation). The fluid commonly selected is a suspension of finely ground magnetite in water. The gravity of separation can be adjusted by varying the concentration of magnetite, providing for greater flexibility than water-only equipment.

Table 1: Specific gravities of coal and its impurities

Material	Specific gravity
Bituminous coal	1.12-1.35
Bone coal	1.35-1.70
Carbonaceous shale	1.60-2.20
Shale	2.00-2.60
Clay	1.80-2.20
Pyrites	4.80-5.20

- Dry separation with air jigs and air tables is generally selected only where liquid-based processes cannot be used—for example, where water is scarce, or a coal does not react well to moisture addition.

Froth flotation is a chemical process used for the recovery of fine coal particles; it does not depend on a difference in density, but instead relies on the surface characteristics of coal and refuse materials.

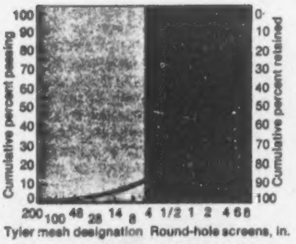
Physical beneficiation in the US is mostly confined to bituminous coals. Lower-rank coals, such as lignite and subbituminous coal, are surface-mined relatively free of impurities. Also, generally they do not wash well because of their high friability and high inherent-moisture content. Cleaning may tend to concentrate sodium in the clean coal product, which increases its slagging/fouling potential (see box on pages S-6-S-7).

Yield and recovery defined

All cleaning processes are judged by their ability to separate clean coal from refuse materials. One indicator used to determine separation efficiency of a process is the percentage of Btus recovered vs the percentage yield or the weight recovered. Yield is expressed as the weight of the clean-coal product divided by the weight of the original raw coal. Recovery, on the other hand, is defined as the Btus recovered in the clean coal divided by the Btu value of the raw coal, multiplied by the percent yield. By definition, cleaning processes maximize yield and recovery at a given quality level. As the level of cleaning deepens, yield falls off much faster than recovery. This produces a low-ash, high-Btu clean-coal product.

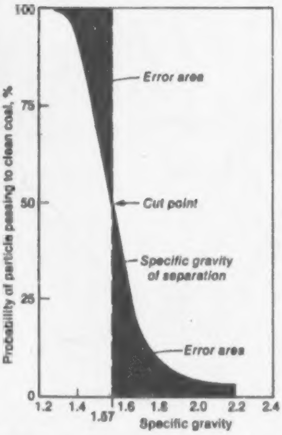
Coal-preparation economics always demand a tradeoff between Btu recovery and ash reduction (coal quality); one is generally improved at the other's expense. By testing coals for washability, it is possible to determine the point at which it is economically justifiable to discard Btus in order to produce a higher-quality coal.

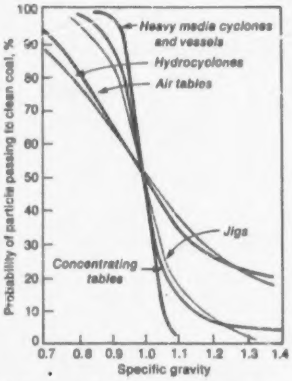
Where to draw the line between clean coal and refuse is purely an economic decision. Under some circumstances, it may be practical to produce two coals having different degrees of cleanliness—a premium product and a middling product (Fig 1). This is often the case with preparation plants for metallurgical coal, which are designed to skim only the highest quality coal from the raw material, leaving behind coal that is perfectly



2. Screen-analysis curve (above) shows percentage by weight of each particle size

3. Performance curves (right) indicate the ability of a process to clean coal





4. Heavy-media systems are the most efficient separators; hydrocyclones least

acceptable for power generation. Middling coal often contains up to 10,000 Btu/lb or more, so, if it is marketed, it can increase recovery considerably and recoup mining and operating costs.

Another case in point is Pennsylvania Electric Co's Homer City preparation plant (Fig 1). This plant is designed to produce (1) a high-quality, low-sulfur coal capable of meeting revised new-source performance standards (NSPS) for SO_2 emissions for combustion in one unit, and (2) a middling product, clean enough for combustion in two existing units that are subject to less rigorous SO_2-emission standards.

The washability study

The ease with which impurities can be separated from a raw coal is known as its washability. Preparation engineers rely on the results of washability studies to determine the correct cleaning-plant equipment and design.

The first stage in a washability study is a screen analysis, to determine the percentage of raw coal particles present in each screen size. Knowledge of the size consistency of coal crushed to a specified top size is necessary to aid the preparation engineer in designing clean-

How cleaning improves the firing characteristics of coal

Many studies have attempted to quantify the effects of coal quality on boiler performance, to justify the use of cleaned coal rather than raw coal. Despite the overriding conclusion that cleaning does reduce boiler-maintenance costs and improve reliability, utility managers still hesitate to buy cleaned coal because its cost/benefit advantages have not been well documented. Thus, an increased understanding of cleaned-coal benefits is in order.

How washing alters coal quality. Poor-quality coal causes slagging, fouling, tube failure, and other boiler-related problems, which reduce efficiency, force deratings, or result in costly unplanned outages. Washing can eliminate some of these problems by reducing the ash and mineral content of raw coal, thereby changing its slagging and fouling potential.

Before discussing how this can be done, it is important to know how the ash in coal is distributed. Coal ash consists of (1) intrinsic ash, amounting to less than 2% of the total, which is bound within the coal, and (2) extraneous ash, consisting of partings of shale and discrete ash particles which can be washed out—depending on particle size and density of the ash and the level of cleaning performed.

There are several advantages of firing coal that has been washed to reduce its ash content:

- Increased pulverizer capacity, reduced wear.
- Decreased slagging potential, increased boiler capacity.
- Precombustion particulate control.
- Decreased capital costs for new boilers and fuel-handling systems.
- Pyrite (sulfur) removal.

Effects of the altered mineral constituency of washed coal are less predictable than those of reduced ash content. In some coals, washing improves slagging/fouling behavior, but the firing characteristics of other coals are unaffected, or even worsened. Washing tends to segregate minerals in ash according to their size/density characteristics, and serves either to enrich or to attenuate those minerals, depending on the washability characteristics of the coal. But washability data cannot be used to determine the slagging/fouling behavior of coal, because they do not distinguish between the various mineral constituents. Instead, a chemical analysis of the ash is the only way to predict the slagging/fouling potential of washed coal.

Washing vs mineral balance. Several studies have addressed the effects of washing on the mineral balance in washed coal. Here are some general conclusions:

- Conventional cleaning of bituminous coals usually reduces the base-to-acid ratio (b/a) and the sulfur content (S), thereby reducing the slagging potential (R_s), where R_s is equal to b/a × S. For example, washing of midwestern high-sulfur coal has been shown to reduce its sulfur content and base-to-acid ratio by 35% and 36%, respectively. The effects of washing on sodium

Table 2: Float-and-sink data for coal sample size ¼-in. by 100 mesh

1		2	3	4	5	6	7	8	9	10
Specific gravity		Dry basis, %			Cumulative results					
					Recovery (float), %			Reject (sink), %		
Sink	Float	Weight	Ash	Sulfur	Weight	Ash	Sulfur	Weight	Ash	Sulfur
—	1.30	35.3	3.35	0.72	35.3	3.35	0.72	100.0	18.42	1.29
1.30	1.35	24.5	6.51	0.92	59.8	4.65	0.80	64.7	26.64	1.60
1.35	1.40	8.2	12.46	1.14	68.0	5.59	0.84	40.2	38.93	2.01
1.40	1.45	4.3	17.25	1.46	72.3	6.29	0.88	32.0	45.68	2.24
1.45	1.50	3.3	22.29	1.74	75.6	6.99	0.92	27.7	50.13	2.36
1.50	1.55	2.9	26.50	1.92	78.5	7.70	0.95	24.4	53.93	2.44
1.55	1.60	2.5	30.71	1.91	81.0	8.40	0.98	21.5	57.57	2.51
1.60	1.70	3.8	36.16	2.74	84.7	9.63	1.06	19.0	61.05	2.59
1.70	1.90	3.6	44.66	3.26	88.4	11.07	1.15	15.3	67.16	2.55
1.90	—	11.6	74.23	2.33	100.0	18.43	1.29	11.7	74.23	2.33

ing circuits with the correct equipment and capacity. In a screen analysis, the coal is filtered through successively smaller screens, from a top size of about 4 in. until all of the fines have been classified. From this analysis, a curve is drawn, showing the percentage by weight of each particle size in the sample (Fig 2).

Float-and-sink test. Washability of the screened coal is determined by float-and-sink tests, in which salt solutions or heavy organic chemicals of varying specific gravity are used to separate high- and low-density coal particles. A series of tanks or separation vessels are filled with liquids of preselected densities, and a quantity of the test sample is introduced into each vessel in turn, beginning with the vessel containing the lowest-density liquid.

After the various particles have stabilized at the top or bottom of the first vessel, the float material is removed and collected on a draining table or on filters, to be dried, weighed, and analyzed. The sink material is removed, drained, and then introduced to the vessel containing liquid of the next-highest specific gravity. This process is continued until all of the solutions have been used. All remaining unfloated material is collected as a final sink product.

The float-and-sink test is normally performed separately on each of the particle sizes selected in the screening analysis, because impurities in different coal types tend to concentrate in different coal-particle sizes.

The choice of test gravities used to make the float-and-sink separations depends on the scope of the project and the amount of information needed. A normal range of specific-gravity selections might include 1.30, 1.40, 1.50, 1.60, and 1.90, with additional intermediate separations if necessary.

Chemical analysis of the specific-gravity fractions within a given size fraction is generally performed to determine the weight of dry coal recovered at each specific gravity, along with percent recovery of ash and sulfur, Btu value, and mineral content (Table 2).

Predicting separation efficiency. Prediction of the percentage yield and recovery from washability data assumes perfect separation at the specific gravity used; however, this is seldom, if ever, possible in practice. No matter what cleaning equipment or process is used, some material—especially near-gravity material—will be misplaced.

oxide (Na_2O) content are variable, causing enrichment in some cases and attenuation in others. The enrichment, however, is usually not enough to offset the accompanying reduction in base-to-acid ratio, and, in most cases, the fouling factor ($R_f = b/a \times Na_2O$) is lowered.

- For lignitic-type ash, conventional cleaning could decrease the base-to-acid ratio, increase the ash-softening temperature, and increase the acid viscosity to reduce slagging. If, however, a large percentage of the sodium oxide in the coal is entrained in intrinsic ash—a common occurrence in western coals—the fouling tendency of the coal could decrease despite intensive cleaning. For example, engineers at Arizona Public Service Co's Four Corners generating station (a mine-mouth plant) attributed increased deratings from slagging to a cleaning process that enriched the sodium content of the ash from 1.81% to 2.19%.
- In wet-bottom units, cleaning could increase the ash-softening temperature enough to initiate slag buildup on heat-transfer surfaces. In other units designed for highly slagging coals, cleaning could reduce slagging sufficiently to cause too much heat absorption in the furnace, and create difficulty in attaining superheat and reheat temperatures.
- In one EPRI study, an analysis of ash by density fractions showed that minerals of low melting temperature were segregated in high-density fractions. Furthermore, all coals tested contained a small fraction of ash high in iron content (specific gravity above 2.8), which is also a slagging/fouling inducer. EPRI showed that ash composition and viscosity characteristics differed among density fractions, indicating that the principles of conventional washing could be used to single out undesirable minerals.

Studies by TVA and DOE. The Tennessee Valley Authority has been one of the most active participants in evaluating coal-cleaning technology. DOE has funded several studies that use TVA plant-maintenance records to amass convincing evidence that clean coal reduces boiler operating and maintenance costs. TVA initiated its investigation because many of the relationships between plant performance parameters and coal characteristics are not clearly defined. By having a better understanding of these relationships, management thought that it would be in a better position (1) to select the optimum type of coal for each specific boiler type in the TVA system, (2) to assess penalties when suppliers fail to meet coal-contract specifications, and (3) to evaluate the cost-effectiveness of coal washing.

DOE and TVA have already begun a 5-7-yr program to quantify the effects of coal quality on performance and reliability. The first stage is an analysis of the maintenance and outage records of TVA coal-fired powerplants. Correlations between coal quality and wear, failure, and frequency of maintenance of components are currently being developed. The second stage involves simultaneous monitoring of coal quality and station performance. Initial tests will be run on a small pilot furnace, then on TVA's Cumberland and Paradise plants, each firing washed coal from recently constructed utility-owned cleaning plants.

Misplacement of coal and refuse is caused by variability in coal-particle shape, and in the fluid-flow characteristics of the cleaning equipment. Partition curves [also called performance, separation, or probable-error (PE) curves] characteristic of a certain cleaning method are used by preparation engineers to determine actual yield and recovery of a cleaning process (Figs 3, 4).

The density of separation, or cut point, is defined as the density of a particle that has a 50% chance of passing to either the clean-coal or the reject pile. A partition curve representing perfect separation is a vertical line passing through the cut point, indicating that all coal lighter than the density of separation reports to the washed product, and all heavier material goes to the refuse steam. The nearer the actual curve approaches this condition, the sharper the separation, and the lower the quantity of misplaced material.

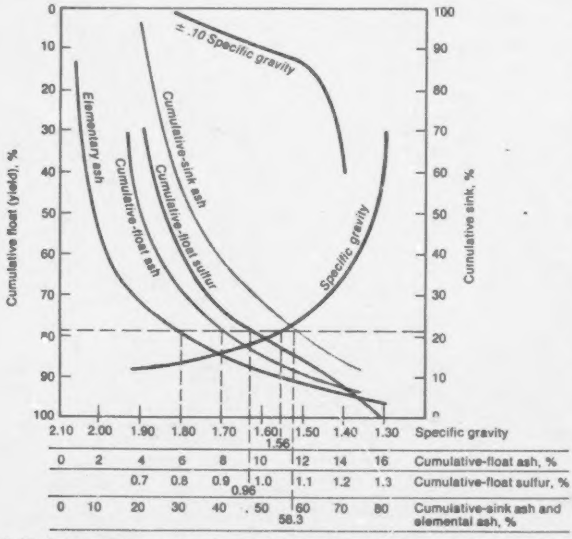
Two methods commonly used to evaluate the separation efficiency of a particular cleaning device: (1) Sum the area between the actual curve and the theoretical curve—the smaller the area, the sharper the separation; and (2) calculate the slope of the curve as it passes through the cut point—a slope of infinity represents perfect separation.

Interpreting washability curves

Using data generated from the washability study, a series of five curves are drawn, with an option for others depending on the requirements of the study. The principal curves include specific gravity (yield), cumulative-float ash, cumulative-sink ash, elementary ash, and ±0.10 specific-gravity distribution (Fig 5). If samples are analyzed for sulfur content, a cumulative-float sulfur curve may also be included.

The specific-gravity (yield) curve is plotted directly from the cumulative-percent (by weight) float data and specific-gravity fractions (column 5 vs column 1 of Table 2). This curve shows the theoretical yield of washed product from the raw coal at any specific gravity of separation. For example, at 1.56 gravity of separation, a 79% yield could be obtained. As the specific gravity is reduced, the yield is also reduced. The steeper the slope, the faster the change in yield with a change in specific gravity. The asymptotic character of this curve illustrates that washing at lower gravities generates more reject material that contains increasing amounts of Btu per unit.

The cumulative-float ash curve is plotted directly from the cumulative-percent (by weight) float and cumulative-percent ash float data (columns 5 and 6), and shows the theoretical percent of ash in the washed product at any given yield of washed product. The flatter the curve, the more the ash that can be removed from the raw coal with the least penalty in yield and recovery. Also, since the specific-gravity and cumulative-float ash curves have common ordinate values, it is possible to determine what the separating gravity would be for a desired ash content.



5. Washability curves, derived from float-sink data, show how well coal can be cleaned

From Fig 6, it can be seen that, to obtain a product having an ash content greater than 8%, the washing gravity should be no greater than 1.56. The maximum possible yield at that separating gravity is 79%.

The cumulative-float sulfur curve is plotted from the cumulative-percent (by weight) float and cumulative-percent sulfur float data (columns 5 and 7), and shows the theoretical percent sulfur in the washed product at any given yield of washed product.

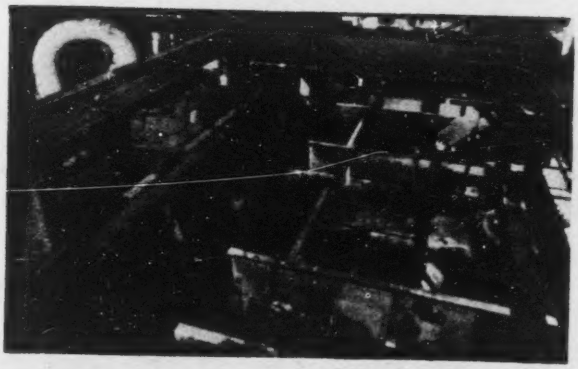
For example, at 79% yield, the sulfur content of the clean-coal fraction would be 0.96%. The flatter the cumulative-float sulfur curve, the greater the sulfur-reduction potential at a specific yield and Btu recovery.

The cumulative-sink ash curve shows the theoretical ash content of the refuse at any yield of washed product (columns 8 and 9). At 79% yield, the refuse would contain 58.3% ash, although it represents only 21% of the raw-coal feed. The flatter the curve, the greater the washability at a specific yield and gravity of separation.

The elementary-ash curve is a derivation of the cumulative-percent ash in the float material, and is intended to show the rate of change of the ash content at different specific gravities or yields. The curve should be smooth unless an error exists in the cumulative-float ash curve. The slope of the elementary-ash curve indicates the relative ease of separation of the coal from the refuse. Steep slopes represent relatively small ash differences for large differences in yield; a shallow slope indicates an easy separation.

The ±0.10 specific-gravity-distribution curve shows the percentage by weight of the coal that lies within ±0.10 specific-gravity units at any given specific gravity. This is the commonly held definition of near-gravity material. It is used as a guide for determining the lowest practical specific gravity to wash a particular coal, and indicates the degree of difficulty of separation. Less than 7% of near-gravity material constitutes simple separation, while separation at a specific gravity with greater than 25% content of near-gravity material is difficult.

The meaning of this distribution curve has changed recently with the advent of heavy-media cleaning equipment that yield high separation efficiencies. While a separation at 25% of near-gravity material would be difficult for a jig or a concentrating table, heavy-media processes can make separations at well over 50% of near-gravity material.



6. Sieve bends are used to dewater coal before it falls to a vibrating screen

Equipment

Physical coal cleaning can require up to four separate material-handling processes depending on the level of cleaning performed. These operations are comminution (size reduction), screening (size classification), concentration (washing), and dewatering.

Comminution

Coal sizing operations at preparation plants are normally confined to crushing. One of the main goals in crushing for a cleaning operation is to achieve a specified top size without creating excessive difficult-to-clean fines. Crushing tailors the coal to the cleaning method used, maximizing efficiency, and liberating impurities that are embedded in or adhered to large pieces of coal. The optimum size to which coal is crushed normally depends on its washability and its end use.

Several types of crushers are used in preparation plants today. They include:

- *Rotary breaker.* This is a large cylinder fitted with radial lifting shelves that lift and drop coal as the unit rotates. It discards stones, shale, logs, and other refuse too large to pass through perforations in the drum. Rotary breakers are commonly used to reduce deep-mined coal containing a high percentage of roof and floor material to a top size between 1 and 6 in.
- *Roll crushers* squeeze coal between tooth-covered roller and stationary breaker plate or between two rolls. These crushers are known for low fines production and are capable of reducing r-o-m coal to 1½ in. or less.
- *Hammermills* throw coal against breaker blocks and grate bars until the product is reduced to the size of the grate opening. Fineness is varied by adjustments in mill speed or by the spacing between hammer tips and grate bars. Hammers can be chosen to permit controlled breaking to produce a minimum of fines.

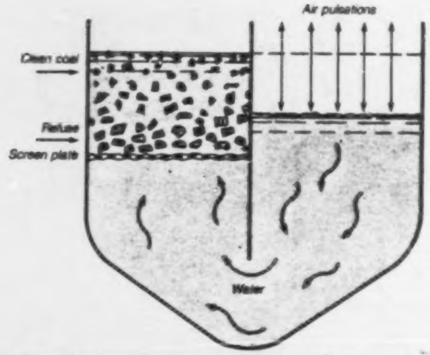
Screening

Screening is necessary to classify coal into different size ranges before introduction to coal-cleaning circuits. Screens also perform a wide variety of functions within the plant. These include:

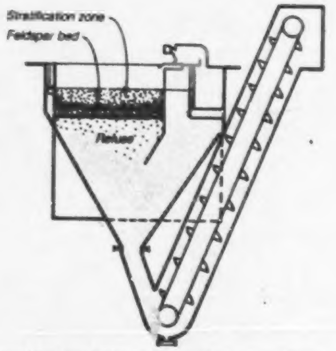
- Removal of rocks and foreign materials prior to crushing.

Table 3: Characteristics of major wet-coal cleaning devices

Device	Medium	Principle of separation	Sizes typically handled	Unit capacities	Specific gravity range	Cleaning efficiency	Comments
Coarse coal jig	Water	Stratification	6 × ¼ in.	High	1.60-1.90	Good	Typically used to clean coarse coal at relatively high specific gravities. Efficiency decreases rapidly at sizes below ¼ in.
Fine coal jig	Water	Stratification	½ in. × 28M	High	1.60-1.90	Good	Reported to be as efficient as concentrating tables when handling compatible size. High unit capacity.
Hydrocyclone	Water	Centrifugal forces in ascending vortex	½ in. × 100M	High	1.60-1.90	Poor	Poor cleaning efficiency, but can be used as 'scalpers' ahead of more efficient devices.
Concentrating table	Water	Stratification and hindered settling	⅜ in. × 200M	Low	1.50-1.80	Good	Provides good separation when near-gravity material is less than 10% of feed. Requires very low water-to-solids ratio.
Heavy-medium vessel	Fine magnetite in suspension	Float and sink	5 × ¼ in.	High	1.30-1.80	Excellent	Efficiency of separation near theoretical, even when feed contains large amounts of near-gravity material.
Heavy-medium cyclone	Fine magnetite in suspension	Float and sink plus centrifugal forces	¾ in. × 28M 1½ in. × 0	High High	1.30-1.80 1.30-1.80	Excellent Very good	Excellent separating efficiency over wide range of specific gravities. Raw feed must be deslimed. Can replace multiple cleaning circuits. Requires no raw-coal desliming.
Flotation	Water	Particle attachment	28M × 0 100M × 0	High Low	N/A N/A	N/A N/A	Not effective for oxidized coal or for removal of pyritic sulfur. More expensive than other cleaning methods.



7. **Jigs use water pulsation** to separate raw coal particles by size and density. These are the most widely used separators



8. **Fine-coal jigs use feldspar beds** to improve stratification of the finer sizes and prevent loss of recoverable fines

- Removal of fine coal prior to crushing.
- Separation of coal into coarse and fine fractions for market or for processing in coarse- and fine-coal circuits.
- Recovery of heavy-media circuits.
- Dewatering of coarse coal.

Wet and dry methods are used for classification. Generally speaking, coal larger than ⅜ in. is capable of being dry screened regardless of its surface-moisture content, whereas classification of smaller sizes requires water sprays to wash fine coal through sieve openings.

Vibrating screens achieve high separation efficiencies by causing stratification by size, inducing fines to settle and sift through screen apertures. Sieve bends are stationary screens with no moving parts and low operating costs (Fig 6). They are used in fine-coal washing circuits, particularly for dewatering.

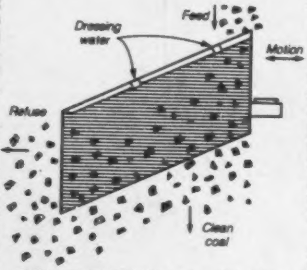
Concentration

This is the step where coal and its impurities are actually separated. The equipment chosen is largely dependent on the coal's size range and washability. A list of the most widely used cleaning devices and their washing characteristics is presented in Table 3. Methods used can be classified as water-only, heavy-medium, and dry separation. Specialized fine-coal recovery methods include froth flotation and oil agglomeration.

Water-only cleaning equipment refers to jigs, concentrating tables, and hydrocyclones. Jigs are among the oldest and simplest of all coal-washing devices and are also the most widely used in the United States, especially for coarse-size coal (Fig 7). Almost half of the coal cleaned in the US in 1978 was done in jigs, although recent emphasis on higher-quality coal and separation at lower specific gravities is causing a progressive shift from water-only devices to heavy-media processes.

In the jigging process, a bed of raw coal placed in a rectangular, open-top, water-filled container is stratified by induced pulsations that rearrange particles according to density. Pulsations (30-60/min) are generated either by air pressure or by a plunger. Heavy refuse material moves to the bottom of the bed (which is generally a perforated screen plate to allow pulse transmission) and the clean coal particles float to the top.

Most jigs are operated to develop three products: a low-density float coal, and intermediate-density middling coal, and a refuse-sink material. The separation efficiency of a jig is determined by the quantity and quality of the middling coal it produces. Because middling coal is made up mostly of near-gravity materials (including particles of quality coal laminated with minerals), this fraction can usually be improved through further liberation and recycling. The tendency in modern plants is to crush the middlings and reclean with heavy-media cyclones.



9. **Concentrating tables** separate by size, shape of particles, as well as by density

Jigging can be applied to a wide size range of coal particles, from a top size of 8 in. down to 28 mesh if the percent near-gravity material is low. For best efficiency, however, small sizes should be washed separately at a different gravity of separation. Reason: The actual gravity of separation increases as particle size decreases because efficient jig operation is dependent on particle size. When a wide size range of coal is cleaned, the efficiency of separation decreases when the particle size decreases. Jigs used to clean only intermediate-size coal (½ in. by 28 mesh) are generally equipped with stone beds to improve stratification of the finer sizes and to prevent loss of recoverable finest to the refuse (Fig 8).

The concentrating table is the most widely used fine-coal cleaning device because of its dependability and efficiency (Fig 9). The units are large tilted rhomboid-shaped decks with ridges (riffles) that span the table surface diagonally. A vibrational drive mechanism is used to create a reciprocating differential motion that causes feed material to fan out onto the deck into strata of different density. The water, which flows against the riffles, is made up partly of feedwater and partly of dressing water issued from the upper side of the table. Clean coal, middlings, and refuse are eventually dispersed so that the finest and heaviest particles concentrate at the bottom of the deck, and the coarsest and lightest congregate near the top. Tabling—one of the gentlest of cleaning processes—has particular application when washing soft and friable coals that degrade easily.

To achieve good separation of coal and refuse, it's important to control variables such as coal size, table shape, feed flow (which should be steady and uniform), and specific gravity of separation.

Hydrocyclones are separating devices that clean medium-to-fine size coal (Fig 10). Key components are a cylindrical and a conical section. Raw coal feed flows tangentially under pressure into the cylindrical section, then descends into the conical section, where particles are subjected to extremely high centrifugal forces. High-density particles (refuse) are pulled outward and downward, and discharge as underflow from the cyclone apex. Low-specific-gravity particles are scooped through a cylindrical vortex finder into the cyclone overflow and discharge as the clean-coal product. Efficient separation of finer-sized particles is often hindered by the resistance of the fluid; as a result, a large percentage of fine coal may slip by as misplaced material in the overflow.

Specific-gravity separation of a hydrocyclone is regulated by varying the diameter of the discharge orifice or by varying the distance between the bottom of the vortex finder and cone bottom. For example, separation gravity can be decreased (less misplaced material to the overflow) by decreasing the vortex finder diameter, increasing the diameter of the underflow orifice, or shortening the length of the vortex finder tube.

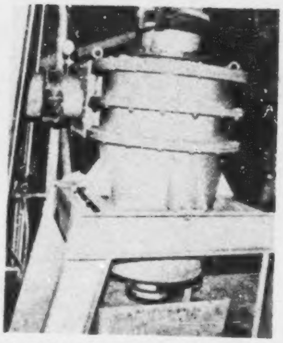
The hydrocyclone is not as efficient a separator as other devices that clean the same size fraction (that is, concentrating tables or heavy-media cyclones). Therefore, two-stage recirculation of overflow material is often recommended to raise its efficiency to acceptable levels (Fig 11). Using this technique, the primary or first-stage hydrocyclone is adjusted to produce an acceptably clean coal, and the secondary or second-stage hydrocyclone is adjusted to produce a refuse that is essentially free of misplaced material. The overflow from the secondary hydrocyclone, containing misplaced coal from the primary cyclone, is returned to the primary-cyclone feed for reprocessing.

Hydrocyclones are often used as scalping units ahead of other beneficiating devices. When used before concentrating tables, their combined efficiency—particularly at low specific gravity—is greater than that of tables alone. Also, because ⅓ to ½ of the feed to the hydrocyclone is usually clean enough to bypass the tables, fewer tables are required. Used ahead of heavy-media cyclones, they reduce the size and cost of preparation and recovery.

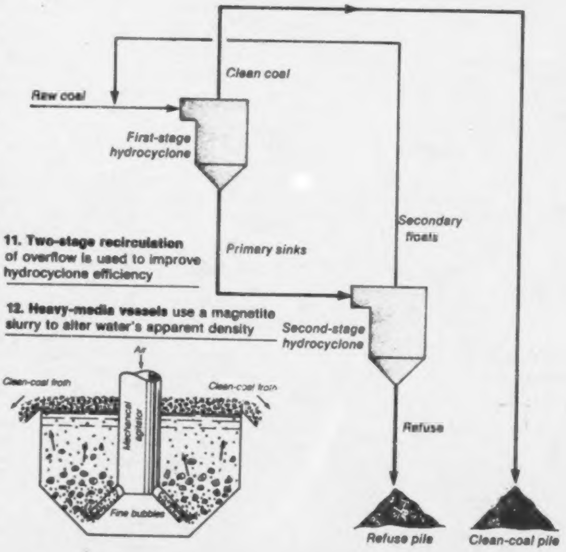
Heavy-media equipment. Cleaning devices based on the principles of heavy-media separation have shown steady growth to the point where over one-third of all coal cleaned today is done either in heavy-media vessels—used to clean coarse coal over ¼ in.—or in heavy-media cyclones, used to clean the remainder.

Heavy-media vessels consist of either drums or troughs containing a dense liquid that could be an organic fluid, dissolved salts, or suspensions of sand, pulverized rock, or, most commonly, magnetite. Magnetite is an iron oxide finely ground to −325 mesh. With a specific gravity of 5.0, it can be used to alter the specific gravity of water solutions.

The heavy-media vessel works as follows: As raw coal is continually introduced at one end of the vessel, particles with a specific gravity greater than that of the medium sink to the bottom and are removed as refuse. Low-density particles float to the surface and are discharged from the vessel as clean coal. Near-gravity materials exit from be-

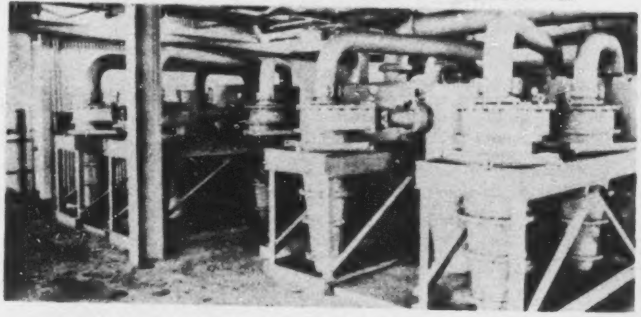


10. Hydrocyclones separate impurities by centrifugal force

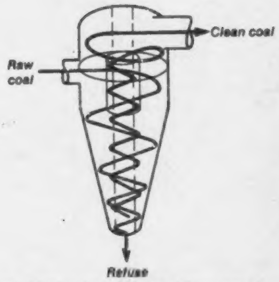


11. Two-stage recirculation of overflow is used to improve hydrocyclone efficiency

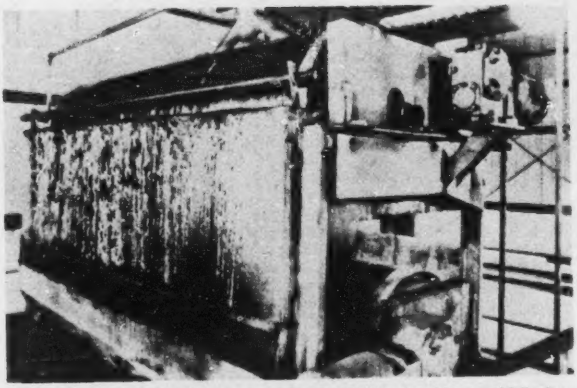
12. Heavy-media vessels use a magnetite slurry to alter water's apparent density



13. Heavy-media cyclones are able to achieve more efficient separation than are hydrocyclones, and are identified by a more acute cone angle than hydrocyclones



14. Cyclone flow path: impurities move to perimeter; coal seeks center



15. This double-roll magnetic separator recovers magnetite rinsed from coal following any heavy media process. Magnetite is then recirculated to the process

tween float and sink material (Fig 13).

Several reasons for the increased use of heavy-media processes are:

- *Flexibility.* Because specific gravity of separation can be altered simply by the addition of magnetite or water, even fluctuations in raw coal quality and changes in clean-coal specifications can be accommodated.
- *Efficiency.* Although heavy-media plants are the most expensive to operate, they are also the most efficient. Sharp separations over a wide range of specific gravities are possible.
- *Versatility.* In recent years, the process has been extended to clean raw coal as coarse as 2 in. all the way down to −28-mesh material. This eliminates the need for multiple cleaning devices.

Heavy-media cyclones (Fig 14) work under the same principle as water-only cyclones, but they are more efficient; they also have a much more acute cone angle and a shorter vortex finder. The flow pattern of media in a cyclone consists of a descending vortex that originates at the inlet and progresses to the underflow outlet (Fig 15). As the media flow down the cyclone perimeter, part of the fluid peels off toward the cone interior, reversing its direction to form an ascending vortex that travels up and out of the overflow orifice.

The specific gravity of the underflow media is greater than the overflow because dense magnetite particles are pulled radially outward by the centrifugal force, which can be as great as 20 g at the entrance and 200 g at the apex. This phenomenon tends to alter the specific gravity of separation to a point higher than the average density.

Dense-media recovery. Drum-type magnetic separators are generally used to recover expensive magnetite from the washing circuit (Fig 16). There are two basic types of separators—concurrent and countercurrent. The magnetite slurry flows with drum rotation in one, and opposite to drum rotation in the other. To cut down on magnetite loss, most manufacturers arrange two magnetic separators in series in the same unit.

The finest magnetite particles are the most difficult to recover because magnetic forces tend to be overpowered as particle diameter decreases. A loss of the finest mesh magnetite could severely influence the sharpness of separation of heavy-media processes, and, if the raw coal feed contains a lot of near-gravity material, the resulting loss in efficiency could be great. Of course, some magnetite is always lost with the clean-coal product or the magnetic-separator tailings, and must be replaced periodically.

To attain the sharpness of separation characteristic of heavy-media systems, specific gravity must be maintained within close limits. Most plants are equipped with automated control systems, which add either water or magnetite to the heavy-media mixing sump whenever required. Some control systems are able to maintain specific gravity within very close limits (±0.003). Generally, the more difficult the separation, the more precise the control required.

Froth flotation has become an economical necessity in most steam-coal beneficiation processes because of rising fuel prices, environmental regulations restricting refuse removal, and increasing fines content of r-o-m coal. Mine owners can no longer afford to toss this valuable fraction to refuse as they did in the past. For this reason, fine coal recovery is one of the most important areas of research in coal preparation today.

Flotation does not rely on the principle of specific gravity separation, but rather on the hydrophobic (water-repelling) or hydrophilic (water-attracting) properties of coal and its associated impurities. Separation is achieved in this method by disseminating minute air bubbles through a feed-coal slurry; the hydrophobic coal is floated to the surface and removed as a concentrate (Fig 17). Tailings (refuse materials) are passed to a static thickener where water is clarified and reused in the plant.

This process requires the use of chemical reagents to establish a water-repelling surface on the material to be floated. Flotation reagents come in three types and are generally used in combinations of two or more. They have the following characteristics:

- Frothers change the surface tension of the water to prolong bubble life.
- Collectors or promoters induce contact between air bubbles and coal particles by forming a water-repellent covering over the material to be floated while selectively ignoring other material.
- Modifying agents enhance the flotation process under special conditions where material selection needs further improvement.

Flotation equipment varies from single-cell to small multiple-cell units. Agitation and aeration are achieved by mechanical, hydraulic, or pneumatic systems. Sharp separation of clean coal and refuse requires sufficient bubble encounter and sustained particle-bubble adhesion. Other factors affecting flotation include particle size, pulp density, coal volatility, reagent type, and dosage control.

One major deficiency of flotation is its difficulty in selectively rejecting pyrites. To eliminate pyrites from the clean-coal concentrate, it is generally necessary to reclean the froth concentrate one or more times using a modifying agent to depress the clean coal. Pyrite is then floated with a normal reagent, skimmed, and discarded.

Oil agglomeration is an alternative to froth flotation and has received wide attention because of the increased emphasis on fines recovery. The principle of agglomeration is similar to flotation: An oil-based liquid is added to raw fines; coal surfaces absorb the oil, and refuse particles repel it. The mix is then agitated so that the oil-covered particles agglomerate into larger particles. Flotation or screening is then used to concentrate the clean-coal product.

Agglomeration is more effective than flotation for particle sizes finer than 200 mesh, and it can also be used to clean coals as coarse as ¼ in. In addition, agglomeration produces a dense, coarse, granular product and does not require vacuum filtering or thermal drying.

Dewatering

Dewatering is necessary after washing to produce a salable clean-coal product and to prepare refuse for disposal. Excessive surface moisture reduces the heating value of the fuel, increases transportation costs, and creates handling and shipping problems. Dewatering is also required to prepare feed for cleaning circuits. The feed to a heavy-media circuit, for example, should be sized, prewetted, and dewatered; fine coal fed to concentrating tables must be prepared to a specified solids-water ratio; and the fine coal fed to heavy-media cyclones must be dewatered before mixing with heavy media.

The method selected to remove water is indirectly influenced by the cleaning process used, since the cleaning process determines particle size. Coarse particles above ¼ in. can be dewatered relatively easily on vibrating screens. Centrifuges are used almost exclusively for size fractions ranging from ⅜ in. to 28 mesh. Anything smaller generally requires a solids-bowl or screen-bowl centrifuge followed by a disk or drum filter.

The following equipment is used to dewater different size fractions:

Screens are classified as either stationary or vibrating, and are used to size, dewater, and deslime (separate the −28-mesh fraction) coal-water mixtures. Vibrating screens are used to dewater coarse-to-medium size coal in preparation for market or for other coal-cleaning processes.

Stationary screens, such as the sieve bend or vortex dewatering sieve (a cross between a centrifuge and sieve bend), are used primarily to dewater fine coal. They may also be used for (1) coarse dewatering of concentrating-table products ahead of vibrating screens and centrifuges, (2) desliming ahead of concentrating tables, and (3) sizing ahead of static thickeners and flotation cells.

Centrifugal driers are of two general types: screen and solid bowl. In both types, centrifugal force of up to 600 G pulls moisture from coal particles as the coal is transported with an oscillating motion to a discharge chute. One of the advantages of the screen-bowl type is that it can handle dilute slurries without hampering performance.

Four types of vacuum filter see service in the coal industry—disc, drum, pressure, and belt. Disc and drum filters are used to dewater suspensions of fine material, either clean coal or refuse, of high solids content.

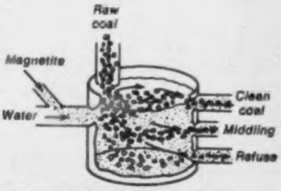
The multiple-disc filter is the most widely used (Fig 18). This device consists of several filter discs up to 15 ft in diameter mounted on a slowly revolving horizontal shaft. A vacuum is applied to each disc while it is submerged in the slurry. This causes the coal to form a cake on the filter medium.

As the discs rotate and the adhesive coal is lifted out of the slurry, a positive pressure is applied to each disc, causing the filter cake to discharge into a receiver. Scrapers are used to dislodge the cake. Drum filters work on the same principle as the discs, but the vacuum is applied to horizontal instead of vertical surfaces.

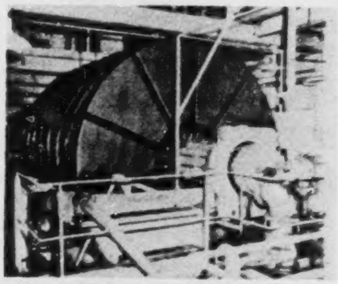
Filter presses achieve much higher moisture removal efficiencies than disc filters, but they are generally not cost-competitive unless the high efficiencies are absolutely necessary. Sludge is pumped between pairs of filter plates between 200 and 300 psi, forcing water out through walls of the plates. The plates are then separated to remove the dewatered product. These devices have been widely used in Europe, but are just beginning to see application in the US.

Belt filters are vacuum and pressure types. In a horizontal-belt vacuum filter, sludge is carried on an endless filter cloth moving over a vacuum chamber until the material is dewatered; then the belt changes direction, and the cake falls off. The double-belt filter uses two endless wire-mesh belts that converge around the sludge and operate to squeeze the liquid out.

Thermal driers. If a coal is unable to meet purchase specifications for moisture content solely by mechanical dewa-

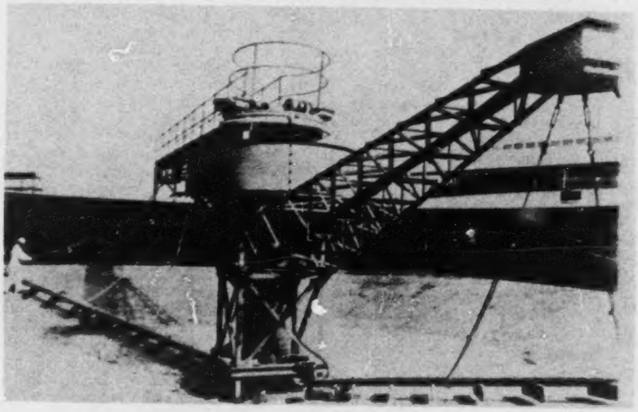


16. Froth-flotation cell uses chemical reagents to make coal adhere to air bubbles and rise to the surface



17. Vacuum disc filters dewater fine coal and refuse of high solids content by sucking material onto huge rotating filters

18. Static thickeners separate fine tailings from spent circuit water by sedimentation

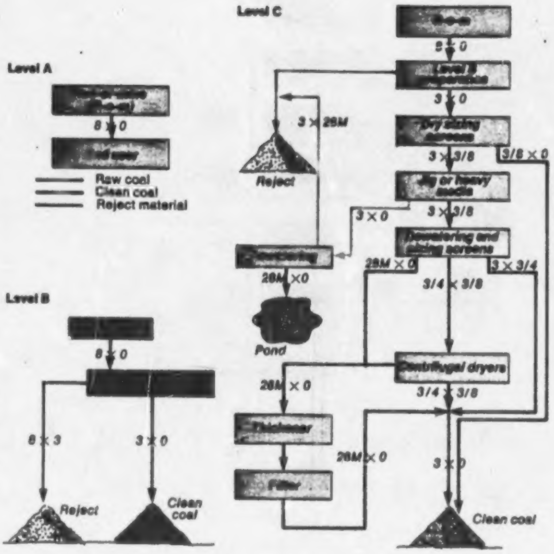


tering methods, thermal drying may be necessary. Drying is also a means of reducing the likelihood of freezeups and of cutting transportation costs. However, it is becoming difficult to justify the cost of this method because of strict particulate-emission-control requirements.

Water clarifiers. Environmental regulations require fully closed water circuits and prohibit the discharge of process water from cleaning processes. All process water must therefore be treated and reused. Since slurry ponds previously used for this purpose have become unacceptable, thickeners are used for slurry sedimentation.

Static thickeners are large pools 50 to 200 ft in diameter, 6 to 16 ft deep, and can hold several million gallons of process water (Fig 19). They process slurries from reject circuits after coarser particles have been removed by mechanical means. The direction of feed to a top center well and dispersion of it through the tank is aided by slow rotation of horizontal rakes, which force settled particles toward an underflow receiving port near the center of the thickener. This underflow travels to a vacuum filter or filter press, where it undergoes further dewatering.

Clarified water is decanted over a circumferential weir along the outside edge of the vessel and is then returned to the plant for reuse. Organic and inorganic flocculants are generally used to enhance separation efficiency.



Systems

The intensity or level of preparation performed on a raw coal depends on its washability characteristics and its market value as clean coal. The advent of strict air-quality standards has created a large market for coals cleaned for sulfur and ash. This trend can be expected to continue as long as cleaning is a less-expensive alternative than FGD as a major best-available control technology. As experience accrues concerning the effects of ash and sulfur on boiler performance, higher levels of cleaning will become more widespread.

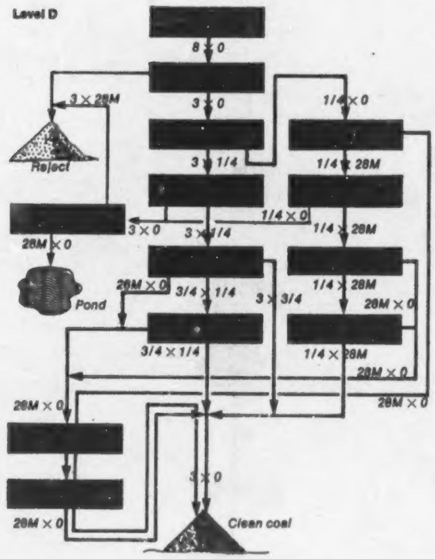
Cleaning plants are classified under the type of circuit used (jig, table, heavy media, etc) and the level of cleaning performed. Most plants fall under one of six basic levels of cleaning (Table 4), which are designed to perform the following operations (also see Figs 18 through 23):

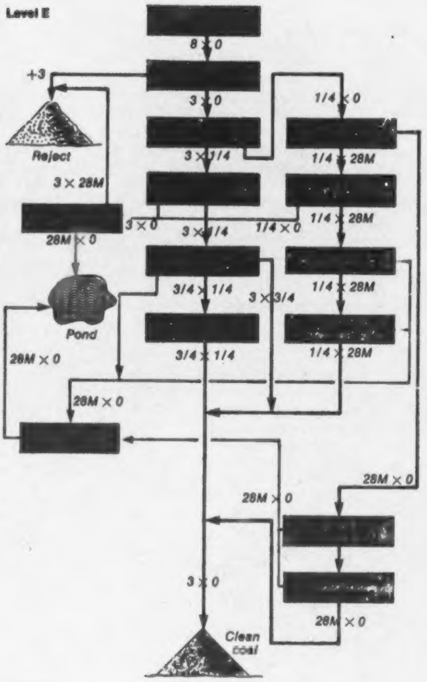
- *Level A.* Absence of preparation. Coal is shipped as mined.
- *Level B.* Breaking for top-size control only, with limited, if any, removal of coarse refuse and trash.
- *Level C.* Coarse beneficiation through washing of ⅜-in. material only; the ⅜ × 0-in. fraction remains dry, and is recombined with clean coal prior to shipment.
- *Level D.* Deliberate beneficiation through washing of all +28-mesh material; the remaining fraction is either dewatered and shipped with clean coal or discarded with the refuse, depending on its quality. Some thermal drying may be necessary.
- *Level E.* Elaborate beneficiation through washing of all size fractions. Thermal drying of ¼ × 0-in. sizes is necessary to limit moisture content.
- *Level F.* Full beneficiation, producing two or more usable coal products from one raw coal in a multistream coal-cleaning system. Outputs are a clean coal conforming to the highest quality standards (generally used for coking), one or more middling products suitable for use as steam coal, and a reject product.

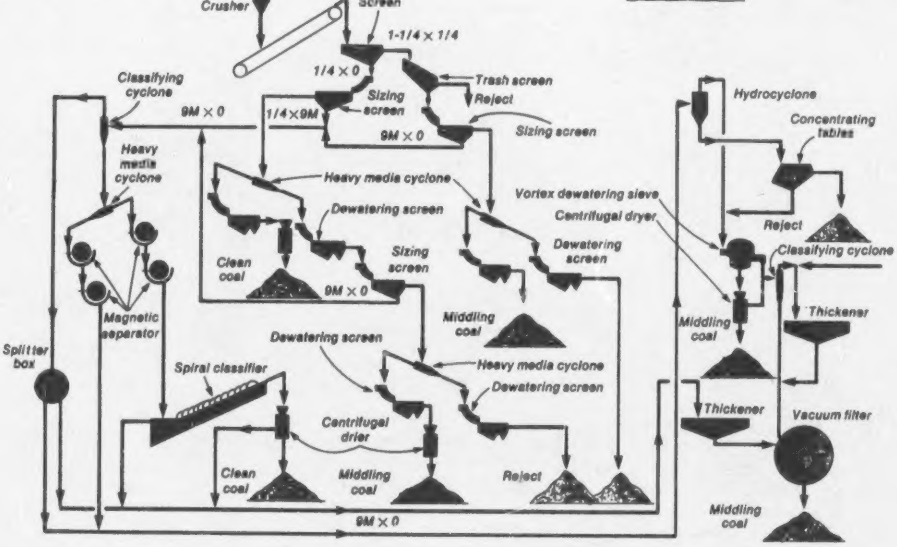
This classification represents an oversimplified explanation of cleaning technology, but is helpful in illustrating and quantifying basic separation principles. Note that each successive level takes into account finer-sized coal using more efficient equipment.

Table 4: Cleaning characteristic of five levels of preparation

Level	Scope	Weight yield %	Btu recovery %	Reduction potential	
				Ash	Sulfur
Absence	No preparation	100	100	None	None
Breaking	Top size control	98-100	100	Fair	None
Coarse	Coarse beneficiation	75-85	90-95	Good	Fair
Deliberate	Fine beneficiation	60-80	80-90	Good	Fair
Elaborate	Very fine beneficiation	60-80	80-90	Excellent	Good
Full	Deep cleaning	50-80	85-95	Adjustable	


Level D
8 × 0
3 × 28M
Reject
3 × 0
1/4 × 0
3 × 1/4
1/4 × 28M
1/4 × 0
3 × 0
3 × 1/4
1/4 × 28M
28M × 0
Pond
28M × 0
3/4 × 1/4
3 × 3/4
1/4 × 28M
28M × 0
3/4 × 1/4
1/4 × 28M
28M × 0
28M × 0
28M × 0
3 × 0
28M × 0
Clean coal


Level E
8 × 0
+3
3 × 0
1/4 × 0
Reject
3 × 1/4
1/4 × 28M
3 × 28M
3 × 0
3 × 1/4
1/4 × 0
1/4 × 28M
28M × 0
Pond
3/4 × 1/4
3 × 3/4
1/4 × 28M
28M × 0
3/4 × 1/4
1/4 × 28M
28M × 0
28M × 0
28M × 0
3 × 0
28M × 0
Clean coal


Level F
R-o-m
8 × 0
Crusher
Screen
1-1/4 × 1/4
1/4 × 0
Trash screen
Reject
Classifying cyclone
9M × 0
1/4×9M
Sizing screen
9M × 0
Sizing screen
Heavy media cyclone
Heavy media cyclone
Dewatering screen
Clean coal
Sizing screen
9M × 0
Magnetic separator
Dewatering screen
Middling coal
Vortex dewatering sieve
Centrifugal dryer
Hydrocyclone
Concentrating tables
Reject
Classifying cyclone
Thickener
Middling coal
Splitter box
Spiral classifier
Dewatering screen
Heavy media cyclone
Dewatering screen
Centrifugal drier
Clean coal
Middling coal
Reject
9M × 0
Thickener
Vacuum filter
Middling coal

24. The new EPRI test facility is the keystone in utility commitment to coal cleaning

Research

Looking down the road to major future developments in coal-cleaning technology, sulfur removal and fines-recovery methods are the center of most industry efforts today. Although such techniques as chemical extraction of sulfur, high-gradient magnetic separation of inorganic minerals, and other new technologies are on the horizon, they are still largely at the pilot stage, and full-scale commercial operation is not expected in the near future because of the costs involved.

The utility industry has just begun to explore cleaning technology for its own benefit. Spearheading utility efforts in this area is the Electric Power Research Institute, R&D arm of the utility industry. EPRI launches its program this month with its own coal-cleaning test facility, located adjacent to the Homer City station of Pennsylvania Electric Co (Fig 24). The full-scale plant is a major project within EPRI's coal-quality program, with announced goals of (1) developing and demonstrating cost-effective cleaning technology to improve fuel quality and uniformity, and (2) quantifying the economic and technical relationships between coal quality, unit performance, and availability.

Test-facility objectives at Homer City are:

- To demonstrate the cleaning characteristics of coals submitted by utilities from around the country, and to provide the washability data necessary for commercial-plant design.
- To advance the state-of-the-art of cleaning equipment, processes, and instrumentation.
- To demonstrate improved sampling and analysis procedures and to maintain a utility-oriented coal-quality data base.
- To provide training for preparation engineers and plant operators.

Many preparation engineers believe the key to bringing the cost of cleaning down is the recovery of valuable fines (smaller than 28 mesh), which have customarily been tossed to the refuse pile because recovery was not practical. Declining r-o-m coal characteristics, increased production costs, and stringent refuse-disposal standards have brought fines recovery to the forefront of cleaning R&D. It is estimated that the Btu recovery of preparation plants, which now averages 80%, could be improved considerably with deeper cleaning.

EPRI plans. With its efforts centered on fines recovery, EPRI plans to investigate these cleaning operations at its test facility:

- Heavy-media cleaning of fines, attempting to minimize magnetite loss and maximize efficiency and control.
- Fines recovery using split-feed flotation (28, 65, and 100 mesh × 0 fractions).
- Washability studies of representative US coals.
- Development of performance models for cleaning and dewatering equipment.
- Degradation of solids related to equipment arrangement and coal type.
- Circuit response to variations in solids concentration.
- Equipment performance and comparison.

The feature that sets the EPRI facility apart from other plants is its ability to accommodate five flow schemes, which can simulate up to 50 different commercial cleaning circuits, based on heavy-media-cyclone, froth-flotation, or concentrating-table operations. The plant is very flexible, and existing equipment can be modified or new equipment added without significant structural changes.

Instrumentation at the test facility will be tied to a central computer capable of performing day-to-day plant operations, such as automatic startup or shutdown of pumps, cleaning and dewatering equipment and ancillary operations, monitoring of all instrumentation, and recording of operating data. Computer-assisted control in coal-cleaning has yet to be demonstrated.

EPRI is also investigating an on-line nuclear analyzing device that will provide immediate feedback of both ash and sulfur characteristics of input and output coal streams. A similar instrument is currently being demonstrated on powerplant fed conveyors at TVA's Kingston plant and Detroit Edison's Monroe plant; however, this instrumentation is still considered experimental.

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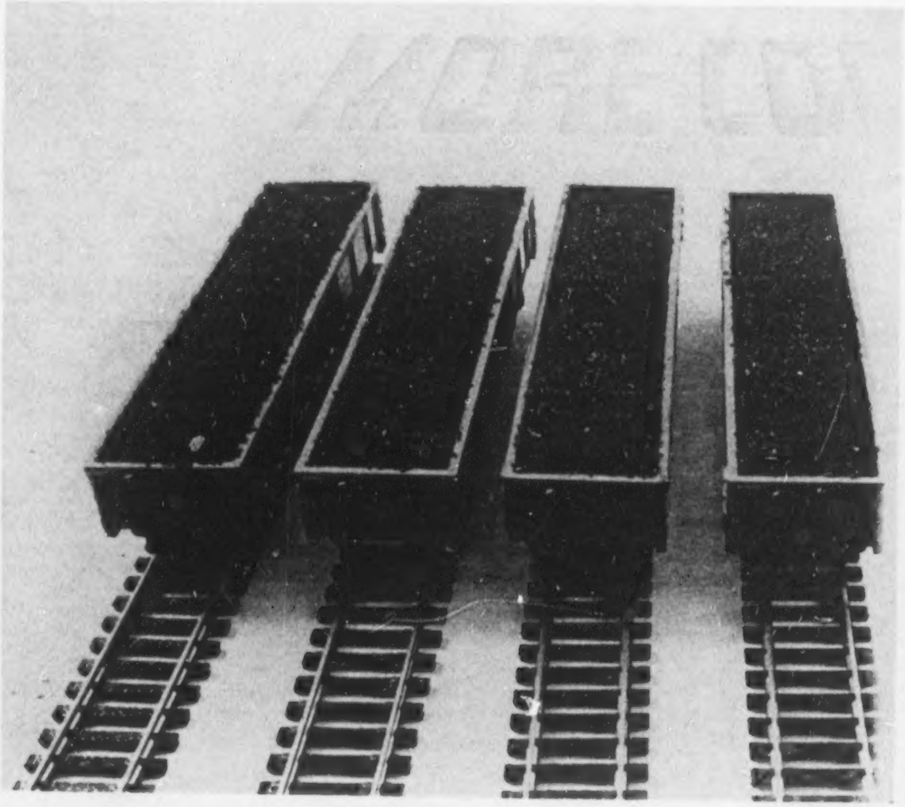
This Electric Power Research Institute article details the positive attributes that result from coal cleaning. "The utility that uses clean coal can benefit from low shipping costs and reduced sulfur emissions and, in some cases, improved boiler availability and reliability. As coal prices rise, as run-of-mine quality diminishes, and as coal use increases, the coal-cleaning option become more attractive."

Most of the coal that electric utilities burn today arrives at the power plant boiler much the same as it left the mine: a less-than-perfect fuel, sullied with sulfur, ash-producing minerals, rock, and moisture. The imperfections make trouble for the utility industry, consumer of two-thirds of the nation's coal output. Rock and moisture bloat shipping costs; sulfur mixes with oxygen during combustion and forms SO_2; minerals give power plant pulverizers a rough time, then form molten ash that can slag and foul boilers and lead to unscheduled outages, deratings, and maintenance difficulties.

Some of coal's imperfections are here to stay, at least for now. Organic sulfur, bound up in the coal itself, cannot be removed, and inherent moisture, present in low-grade coals, is difficult to eliminate. However, many impurities can be reduced through physical coal cleaning: ash-producing minerals (ash) and pyritic (inorganic) sulfur can be minimized by existing crushing and separation processes; surface moisture can be evaporated. This results in a coal with improved Btu content—an all-round better fuel.

Changing circumstances

Despite the benefits of clean coal—lower shipping costs, improved boiler operation, reduced sulfur emissions—only about 20% of the nearly 500 million tons



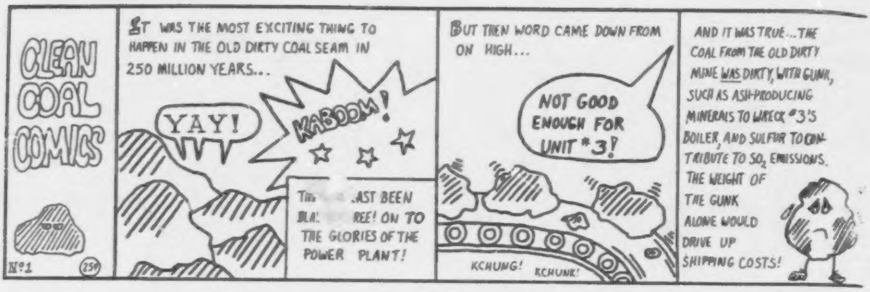
used yearly by the utility industry is cleaned, and that minimally. The rest is used raw. The reason is economics. "All things considered, raw coal has been cheaper," maintains Shelton Ehrlich, manager of EPRI's Fluidized Combustion and Coal Cleaning Program. Capital costs for coal-cleaning plants soar from $3000 to $31,000 per ton per hour of installed capacity, depending on the level of cleaning. Actual operating costs at these plants run from 80¢ to $4 per ton of raw feed. These costs are passed on to the utility industry by mine operators, who do most of today's coal cleaning. And so far, the benefits of clean coal to utilities just haven't balanced the high cost of cleaning.

A tangle of developing circumstances—including higher coal prices, diminishing coal quality, increasing coal use, and mounting environmental awareness—are tipping the balance in favor of coal cleaning in many cases. Suddenly, present-day cleaning techniques are undergoing careful scrutiny by interested parties, and where those techniques are found wanting, research (such as EPRI's $25 million coal cleaning effort) is being intensified to increase efficiency and lower costs.

Past prices

A look at past coal prices should begin to unravel today's coal-cleaning economics. The average price of raw, deep-

Clean Coal. From four carloads of raw coal, roughly one carload of impurities can be removed by physical coal cleaning. The utility that uses clean coal can benefit from lower shipping costs and reduced sulfur emissions, and, in some cases, improved boiler availability and reliability. As coal prices rise, as run-of-mine quality diminishes, and as coal use increases, the coal-cleaning option becomes more attractive.


CLEAN COAL COMICS
Nº1
25¢
IT WAS THE MOST EXCITING THING TO HAPPEN IN THE OLD DIRTY COAL SEAM IN 250 MILLION YEARS...
YAY!
KABOOM!
TH[illegible]AST BEEN BLA[illegible]REE! ON TO THE GLORIES OF THE POWER PLANT!
BUT THEN WORD CAME DOWN FROM ON HIGH...
NOT GOOD ENOUGH FOR UNIT #3!
KCHUNG!
KCHUNK!
AND IT WAS TRUE...THE COAL FROM THE OLD DIRTY MINE WAS DIRTY, WITH GUNK, SUCH AS ASH-PRODUCING MINERALS TO WRECK #3'S BOILER, AND SULFUR TO CONTRIBUTE TO SO2 EMISSIONS. THE WEIGHT OF THE GUNK ALONE WOULD DRIVE UP SHIPPING COSTS!

mined coal no more than 10 years ago was $3–$5 a ton, and at those rock-bottom rates, utilities found it easy to forgive coal's faults. Rather than pay top dollar for premium or cleaned coal, the industry preferred to pay low prices for raw coal, absorbing the cost of shipping superfluous rock and moisture and compensating for the fuel's shortcomings by conservative boiler designs. Emissions standards were still in their infancy. All that utilities really needed was heat to generate the steam to generate electricity, so they bought their fuel on a cents-per-million-Btu basis, according to Kenneth Clifford, manager of EPRI's coal-cleaning research. Accordingly, coal quality specifications for the utility industry have been minimal.

Other coal users are more exacting in their requirements. The steel industry, in particular, must specify narrow quality parameters for coking coal to ensure the production of high-quality steel. Consequently, steelmakers pay premium prices for top-quality coking coal, most of which is cleaned at preparation plants.

Today, some utilities buy cleaned coal to satisfy state or local air pollution restrictions, explains Ehrlich. Others, those that hold captive mines rather than having to buy on the market, clean their own coal and can benefit from lower shipping costs. Still, what cleaning is done is minimal, performed at relatively simple preparation plants.

Present situation

Mining costs are rising, and so is the price of coal. A ton of deep-mined subbituminous coal commands anywhere from $25 to $40 today; a ton of strip-mined coal, $10–$15. Coal is the single highest cost component of a coal-fired power plant's operating and maintenance expenses.

At the same time, general run-of-mine coal quality is going down. The nation's prime coal seams are being depleted through aggressive mining, and the use of mechanized and continuous-mining methods (which produce a poorer-quality product) is increasing.

Coal use is on the upswing as the nation's other fuel sources are choked off. Utility coal consumption in 1978 amounted to 480 million tons. By 1985 the figure is expected to reach 800 million tons, and by 2000, about 1600 million tons.

Combine increasing coal use, higher coal prices, and diminishing run-of-mine quality with the possibility of lower shipping costs, better boiler operation, and reduced emissions, and coal cleaning begins to look much more attractive.

Cleaning may eliminate waste that accounts for perhaps 15–35% of the raw coal's mass, paring away at shipping expenses.

Coal cleaning can also remove up to 65% of the ash, resulting in improved boiler availability and reliability, especially with older boilers. Most older boilers are less tolerant of higher-than-design ash coals and are more prone to ash-related fouling, slagging, and corrosion problems, with attendant deratings, unscheduled outages, and maintenance difficulties. Newer boilers are, or can be, designed to be more trouble-free, although they are still subject to derating, observes John Dimmer, a project manager with EPRI's Fossil Plant Performance and Reliability Program.

The effects of ash on power plant operations have yet to be precisely quantified and may depend much on site-specific factors, cautions Dimmer. But by and large, higher ash content usually spells trouble for older boiler systems, and cleaning the ash out of coal before combustion may prove a boon to many troubled utilities. Reduced ash content also reduces the load on electrostatic precipitators and baghouses, whose job is to remove fly ash from power plant emissions. Less ash in coal also means that the power plant operator has a smaller volume of combustion by-products to dispose of, an important consideration in view of possible EPA regulations affecting coal ash disposal.

Coal-cleaning credit

The final coal-cleaning incentive for electric utilities is the chance to eliminate some of coal's inorganic sulfur content before combustion can transform it into

DEJECTED, THEY WAITED IN A STOCKPILE FOR WEEKS...
AND WEEKS...
UNTIL, ONE DAY...
CLEAN 'EM UP!!
THE ORDEAL BEGAN WITH SIZE REDUCTION—BREAKERS AND CRUSHERS KNOCKED OFF THE LARGEST IMPURITIES AND EXPOSED HIDDEN ONES...
LUMPS ON MY LUMPS!

SO_2, necessitating a costly postcombustion scrubbing with flue gas desulfurization (FGD) equipment. As much as 30–35% of coal's total sulfur content could be removed by cleaning, according to Ehrlich, if technological improvements in the coal-cleaning process can be made. Costly FGD scrubbers would still be necessary to meet the June 1979 revised New Source Performance Standards. The law limits sulfur dioxide emissions to 0.6–1.2 pounds or less per million Btu of coal fired and sets a sliding scale of 70–90% sulfur removal. A coal-cleaning credit is included in the new regulations. The utility that augments scrubbers with clean coal processes stands to save money on FGD-related costs, such as sludge disposal and limestone. Exactly how much clean coal can save over scrubbing is now the subject of an EPRI study, but Ehrlich guesses, "It sometimes costs 10 times as much money to remove a pound of sulfur with scrubbers as it does with coal cleaning."

The credit will be a powerful incentive for coal cleaning, maintains Ehrlich. In addition, the high sulfur content of many U.S. interior and Appalachian coals may make them unusable under the new EPA standards, even with scrubbing, unless the coal is vigorously cleaned, according to Kurt Yeager, director of EPRI's Fossil Fuel Power Plants Department.

A coal-cleaning credit might also offer utilities in some areas of the country a measure of flexibility in responding to emission regulations, suggests Stephen Baruch, EPRI technical manager for Environmental Affairs. Suppose, for example, a utility burns raw coal and requires more generation, yet is in an environmentally sensitive area and just meets emissions standards. The utility might switch from raw coal to clean coal, and the resulting lower emissions could provide just enough leeway to allow installation of needed generation capacity. Similarly, a utility using raw coal and still meeting emissions regulations in an otherwise sensitive area might be able to shift to clean coal and sell or bank its unneeded credit.

From art to science

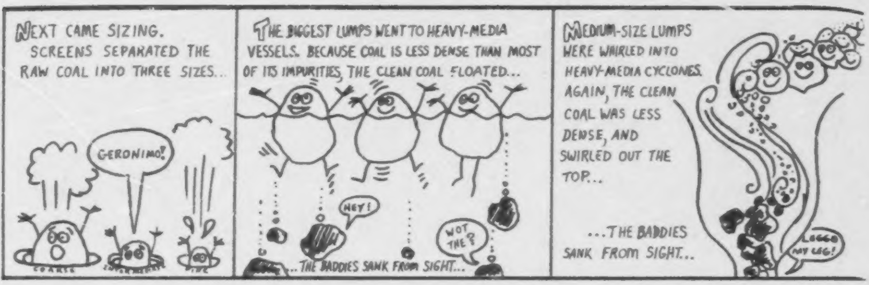
Coal preparation—a broad term that encompasses all coal processing from mine to power plant, including coal cleaning—is of significant potential value to the nation's electric utilities. But the state of the art today is exactly that, an art. Preparation must now be transformed from an art into a science, asserts Randhir Sehgal, EPRI project manager in coal-cleaning research. Existing technologies, some admittedly primitive, must be improved upon, and other, newer, technologies developed.

Coal preparation engineers pursue two major goals: improving removal of impurities—ash-producing minerals, sulfur, and moisture—and reducing Btu losses.

Pyritic sulfur and most ash-producing minerals either happened to share the same swamp as the plants that were metamorphosed into coal or were deposited into the coal bed at a later date. Also, shale, slate, sandstone, and limestone from the roof and floor of the mine are often extracted with the coal. These extraneous impurities can be removed by physical crushing and separation, and the only limits on how much can be removed are economics and technology. (Some pyrites, for example, are so finely and evenly disseminated throughout coal that removal is virtually impossible.)

The organic sulfur, as well as small amounts of ash-producing minerals, is chemically combined with the coal. These inherent impurities cannot be removed by physical crushing and separation. Chemical coal-cleaning methods, such as leaching with acid, can remove a small percentage of organic sulfur; more advanced methods, such as solvent refining (which actually dissolves the coal under increased temperature and pressure), would eliminate more of this sulfur. However, such chemical methods more than double the cost of coal, according to Ehrlich, and have not yet been commercialized.

Total ash content in run-of-mine coals can typically range from as low as 9% to as high as 30%. Total sulfur content, often roughly half pyritic, half organic,


NEXT CAME SIZING. SCREENS SEPARATED THE RAW COAL INTO THREE SIZES...
GERONIMO!
COARSE
THE BIGGEST LUMPS WENT TO HEAVY-MEDIA VESSELS. BECAUSE COAL IS LESS DENSE THAN MOST OF ITS IMPURITIES, THE CLEAN COAL FLOATED...
HEY!
WOT THE?
...THE BADDIES SANK FROM SIGHT...
MEDIUM-SIZE LUMPS WERE WHIRLED INTO HEAVY-MEDIA CYCLONES. AGAIN, THE CLEAN COAL WAS LESS DENSE, AND SWIRLED OUT THE TOP...
...THE BADDIES SANK FROM SIGHT...
LEGGO MY LEG!

varies from less than 1% to over 6%. Inherent moisture, which lowers a coal's fuel value and adds to shipping costs, can fluctuate from 1% in top-quality coal to over 45% in low-grade coal. It is extremely difficult to remove. Surface moisture, on the other hand, is more easily removed.

Once ash and sulfur have been removed from raw coal, they must be properly disposed of to avert possible environmental problems, reminds Stephen Baruch.

In the process of mining, handling, and cleaning, as much as 25–30% of the coal is reduced from coarse lumps to fine particles (called fines). These fines are difficult to clean and expensive to dewater. They can also be a dust problem and a fire hazard. At many plants, even today, fines are simply discarded. Despite the nation's much-vaunted wealth of coal reserves, those lost fines represent lost Btu and lost money. As Ehrlich is quick to point out, "Coal is worth nothing underground. When you go to the trouble of proving a reserve, the coal is worth 10¢ a ton. By the time you've mined the coal and loaded it into a railroad car, it's worth $25 a ton." Ehrlich concludes, "It's unacceptable to leave the fines, perhaps 25% of this product, in a waste pond or refuse pile." Coal preparation engineers are zeroing in on improved means to recover these valuable fines.

The plant itself

There are some 500 coal preparation plants in the United States, typically designed when coal prices were low in relation to capital cost and substantial Btu losses were acceptable. These plants use a variety of circuits, or systems. Some simply screen and crush coal. Others clean the coarse coal and mix it with the raw fines before shipment. Still others clean unsized feeds. Modern plants separate coal into as many as four size ranges, or fractions, cleaning each fraction with the equipment best suited to it.

There are also advanced physical cleaning plants. For example, at Homer City, Pennsylvania, a 1200-t/h facility is undergoing testing by its owners, General Public Utilities and New York State Electric and Gas, cofunded by EPRI and EPA. EPRI is sponsoring the design, construction, and initial two-year operation of a flexible 4–20-t/h coal-cleaning test facility near Indiana, Pennsylvania. Construction of this $6 million installation is expected to begin this summer, and completion is scheduled for late 1980. Technology development under EPRI management will begin shortly thereafter.

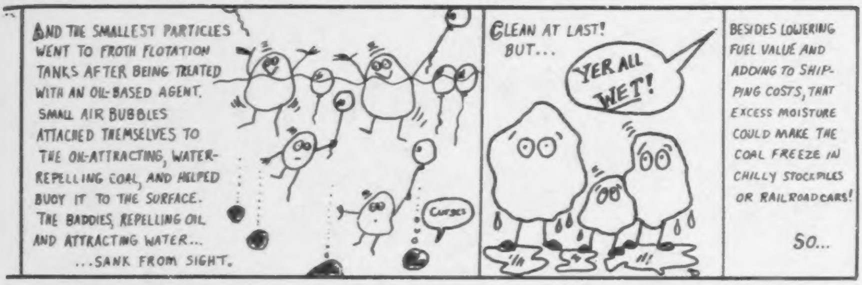
Although preparation specifics vary from plant to plant, the initial treatment is much the same everywhere: when run-of-mine coal arrives at the preparation plant, it first undergoes size reduction, or comminution. This process, largely accomplished by breakers, liberates coal from grosser rock and debris. These devices essentially drop the raw lumps so that the softer coal shatters while the harder rock remains intact and is discarded. For very hard coals, a crusher is necessary. Breakers and crushers both do the jobs they are supposed to do, but perhaps too well: the fines produced, particularly in crushers, are excessive, and better comminution techniques are needed to minimize fines production.

After size reduction, plant flowsheets begin to diverge. Some clean unsized feeds; others separate coal into different size fractions with the help of a series of dry and wet vibrating screens, and clean each fraction separately. Modern plants typically treat three coal fractions: coarse, intermediate, and fine.

All cleaning circuits in use today operate on physical principles, and most operate on the gravity separation principle. Gravity separation takes advantage of the fact that coal has a lower specific gravity than do most of its impurities, and, therefore, the impurities settle out at a faster rate than coal does.

Coarse-coal cleaning

One popular device for cleaning coarse coal is the jig, although it may be used on a full range of coal sizes. In a jig, coal rests on a screen plate. Currents of water are pumped up through the plate, and when the feed is pulsated by the moving water, the less-dense particles—coal—


AND THE SMALLEST PARTICLES WENT TO FROTH FLOTATION TANKS AFTER BEING TREATED WITH AN OIL-BASED AGENT. SMALL AIR BUBBLES ATTACHED THEMSELVES TO THE OIL-ATTRACTING, WATER-REPELLING COAL, AND HELPED BUOY IT TO THE SURFACE. THE BADDIES, REPELLING OIL AND ATTRACTING WATER...
...SANK FROM SIGHT.
Curses
CLEAN AT LAST! BUT...
YER ALL WET!
BESIDES LOWERING FUEL VALUE AND ADDING TO SHIPPING COSTS, THAT EXCESS MOISTURE COULD MAKE THE COAL FREEZE IN CHILLY STOCKPILES OR RAILROAD CARS!
SO...

move toward the top of the mixture. The more-dense particles—refuse—tend to sink. This process may be repeated as many as 50 times a minute.

Jigs have a drawback, however. When they must clean a feed that contains a large amount of near-gravity material (i.e., refuse whose specific gravity is very close to the coal's specific gravity) their efficiency goes down.

The heavy-media vessel and the heavy-media cyclone are two other devices used for cleaning coarse coal. They are the most efficient systems of coal cleaning presently available, functioning well in the presence of large amounts of near-gravity material.

Heavy-media, or dense-media, cleaning works by gravity separation. The heavy-media vessel is a tank or drumlike device through which move streams of raw coal and a heavy medium, usually magnetite and water. The specific gravity of the heavy medium is the gravity of separation, between what the engineer considers coal and what he considers refuse. As the raw coal slowly travels through the medium, the more-dense refuse particles settle out and leave the vessel through a low-level exit; the less-dense coal particles exit at a higher level. The magnetite is later recovered from both clean and refuse fractions with magnetic separators.

The heavy-media cyclone has the added benefit of centrifugal action. Cleaning is done in a cone rather than a tank. Coal and the medium are whirled into a stationary cone; the combination of centrifugal force and gravity does the work. The more-dense refuse tends to concentrate at the sides of the cone and exits at the bottom; the less-dense coal swirls up and out an opening in the top of the cone.

Separation gravities in heavy-media systems can be controlled by adding or removing magnetite. However, these systems are generally costly to install and operate, and they do not work well on fine coal when an entire range of coal sizes is fed to the unit.

Intermediate-coal cleaning

Intermediate-size coal fractions may be cleaned by heavy-media cyclones, as well as by water cyclones and cleaning tables.

Water cyclones, or hydrocyclones, are essentially the same as heavy-media cyclones, except that water rather than a magnetite slurry is the separating medium. A coal-water slurry is pumped into the cyclone; the less-dense clean coal exits through the top of the cyclone, and the more-dense refuse leaves through an opening in the bottom. Like most water-based coal-cleaning processes, water cyclones operate inefficiently when the slurry contains large amounts of near-gravity material.

Cleaning tables (also called concentrating tables) are slightly tilted, ribbed, vibrating surfaces. A coal-water slurry is fed in at one corner of the top of the table; as the slurry moves down across the table, the mixture fans out in a predictable pattern: the more-dense refuse moves toward the bottom of the table, the less-dense, clean coal remains nearer the top. Cleaning tables are relatively inexpensive cleaning devices, but, like the water cyclone, they are inefficient performers when faced with sizable amounts of near-gravity material.

Fine-coal cleaning

The most difficult of all coal fractions to clean are the fines. As coal particles get smaller, specific gravity differences become less and less significant. Many plants clean fines along with coarser sizes in gravity separation equipment, but since a fine-coal particle behaves much the same in solution as a fine-refuse particle, low fines-cleaning efficiencies are the result.

Froth flotation, which relies on surface properties rather than on specific gravities to make a separation, is a common method presently used on fines. In essence, coal attracts oil, repels water; refuse repels oil, attracts water. So raw fines are treated with an oil-based collecting agent that adheres to the coal particles rather than to the refuse. The fines slurry is then fed into a tank, or froth flotation cell. Mechanical agitation

SOME COAL WAS DRIED ON VIBRATING SCREENS...
SOME IN CENTRIFUGES...
SOME BY THERMAL DRYERS.
FINALLY, CLEAN AND DRY, THE COAL FROM OLD DIRTY PASSED MUSTER...
LOOKIN' GOOD!
GRADE A ILLINOIS
LOW SULFUR
LOW ASH
UNIT #3 — HERE WE COME!
The End!

induces air into the cell, and the small air bubbles attach themselves to the oil-coated coal. The bubbles rise to the surface in a froth, taking the clean coal with them. The refuse remains in the slurry, and is rejected as tailings. Unfortunately, froth flotation is not perfect: Fine pyrite particles tend to repel water, so a considerable percentage rises with the clean coal; fine tailings are a growing disposal problem; accurate instrumentation and control are needed.

Another means of cleaning fines (still under development) is oil agglomeration. An oil-based liquid is added to a mixture of fine raw coal. The coal surfaces absorb the oil; the refuse surfaces repel it. The mix is then violently agitated so that the oil-covered coal particles agglomerate into larger particles. These can then be collected by flotation or screening.

After cleanup

Once coal has been through the gamut of cleaning equipment at a preparation plant, it's soaking wet. Besides lowering fuel value and adding to shipping costs, this excess surface moisture may permit coal to freeze in stockpiles or railroad cars, creating handling problems. Plant operators use dewatering methods to remove much of this surface moisture. Vibrating screens are usually adequate for dewatering coarse coal; centrifuges and filters are used for intermediate and fine sizes. Occasionally, thermal dewatering with fluidized-bed dryers is necessary. However, neither mechanical nor thermal dewatering works well with fine coals, and improvements are needed.

Dry cleaning of coal is one way to get around dewatering problems. A number of dry cleaning methods are already in limited commercial use, including the air jig, which performs gravity separation by using pulsating currents of air. Others, such as magnetic separation, electrostatic separation, and the use of heavy organic liquids as heavy media, are still under development. However, dry systems generally suffer from poor efficiencies.

Instrumentation and automation of cleaning plants are needed to cut costs, increase output, and maintain product consistency. There is little instrumentation or automation at preparation plants today, according to Sehgal. Instead, most plant and unit process control is provided by the skill of individual operators. Equipment that could monitor the fluctuations of feed size, moisture, ash, and sulfur would contribute to improved plant efficiency.

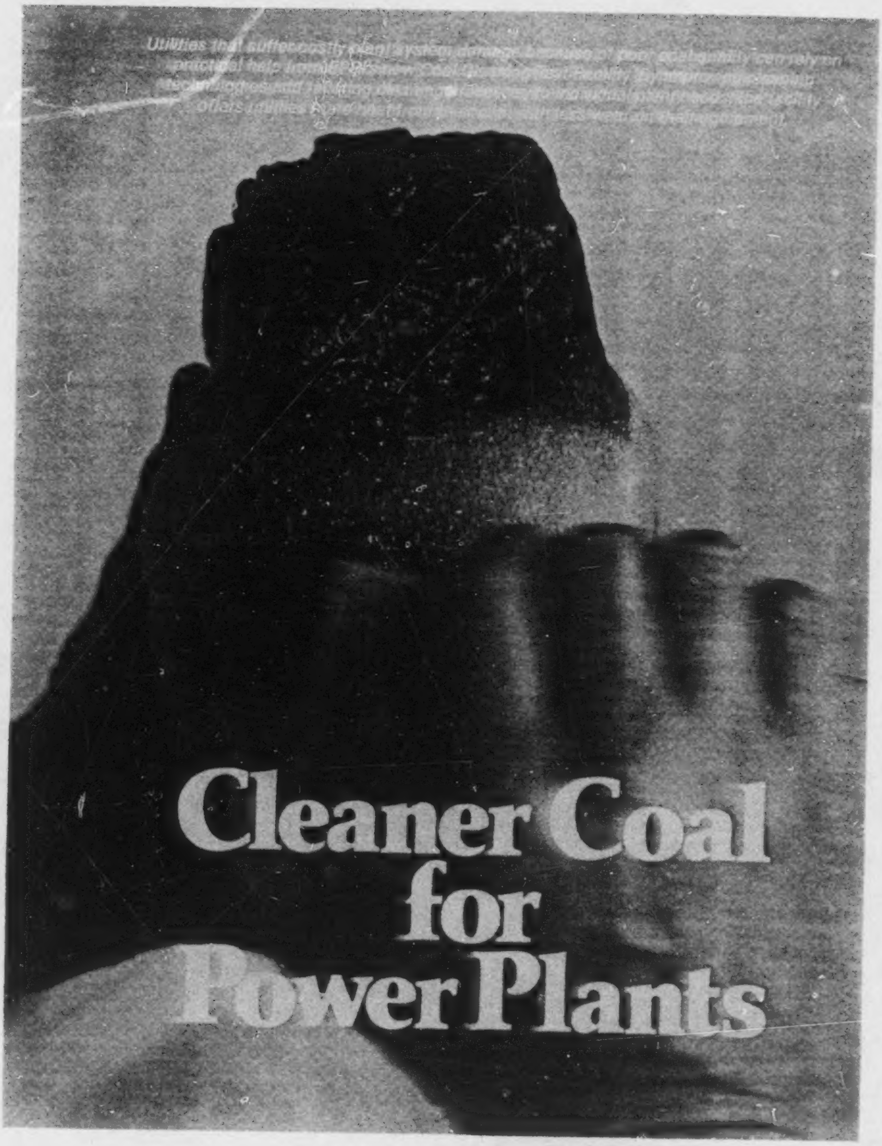
Automation, including control of feed rates according to feed characteristics and control of reagent additions to froth flotation systems, would improve the situation further.

Another area that requires improvement is preparation plant design. Engineers must often rely on insufficient data when they design new plants: either exploratory coal samples are too small to represent actual mine conditions or the effects of mining are not properly simulated in trial washability tests. As a result, operators are frequently in for what Sehgal terms "many nasty surprises" when a new plant goes on-line, and better analysis and sampling techniques are needed to ensure plant efficiency.

The ultimate beneficiaries

The R&D necessary to achieve better ash and sulfur rejection and increased Btu yields is sizable, and not cheap. But if better ways to clean coal can be developed with only small increases in capital and operating costs, the potential benefits to utilities are no less sizable. Better ash and sulfur rejection leads to better fuel, lower shipping costs, improved boiler reliability and availability, and emission control alternatives. Increased Btu yields lead to lower mining costs, and because coal is sold on a highly competitive, buyer's market basis, asserts Ehrlich, these savings will be promptly reflected in lower coal prices. Utilities and utility customers are the ultimate beneficiaries of coal cleaning, and that, concludes Ehrlich, is why the industry must hone this art to a science.

Electric Research Power Institute (ERPI) has designed a facility near Homer City, Pennsylvania, to test various U.S. coals and determine the most economical way to clean them. EPRI encourages coal cleaning for the economic benefits which can accrue, including reduced requirements for add-on environmental controls, decreased maintenance problems and greater energy yields.



Utilities that suffer costly plant system damage because of poor coal quality can rely on practical help from EPRI [illegible]
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Cleaner Coal for Power Plants

Minerals in the primeval vegetation that decomposed to form coal are still contained in coal deposits, along with other minerals left by water flows over the eons. A higher percentage of these minerals is showing up in mined coal. In fact, over the last 10 years, this greater mineral content has meant that the average quality of coal burned in power plants has fallen in terms of its energy-producing potential from 11,750 Btu/lb to 10,600 Btu/lb. Over the same period, the availability of U.S. electric generating plants has dropped by an average of about 10% because of a steady rise in plant outages.

These parallel phenomena are often believed to be related; consequently, many utilities are investigating the characteristics of their coals. Coal quality in general has deteriorated because of changes in mining methods and regulations. Mine productivity is down, and this decline has hastened the spread of mechanized mining methods, which discriminate less precisely between coal and mine roof and floor.

Although power plant boilers can be designed to burn almost any grade of coal, the firing of coal that is inferior to that specified results in an increased outage rate. Many boilers on-line were designed for higher-quality coal than they are now firing, and researchers are trying to determine how much the performance of these boilers would improve if coal quality were upgraded by coal cleaning to the level originally specified. Although several hundred commercial coal-cleaning plants exist, their processes cannot separate all the Btu (energy-producing content) from the mineral matter, and therefore about 5% of the Btu are disposed of with the refuse.

Five years ago, EPRI saw the urgent need for a scientific approach to upgrading the quality of coal and coal cleaning for power plants. After consulting numerous experts in the field, EPRI began to plan construction of an advanced coal-cleaning test facility. The project had three objectives: to characterize steam-coal cleanability on a national basis; to develop and test new equipment and processes; and to train coal-preparation engineers and operators.

The $15.2 million facility was completed last year near Homer City, Pennsylvania, and is being operated by Kaiser Engineers, Inc. Cosponsoring the facility with EPRI are Pennsylvania Electric Co., New York State Electric & Gas Corp., and the Empire State Electric Energy Research Corp. Through services offered by the facility, electric utilities can find out which cleaning processes are the most effective for their particular coal.

Generation costs

Even though the purchase of coal incurs the largest chunk of the operating expenses at a coal-fired plant, the cheapest coal does not always result in the lowest-cost electricity. Low-cost, high-sulfur coal, for instance, can be a financial liability as far as controlling stack emissions is concerned.

Most coal delivered to utilities contains impurities. These include certain minerals inherent in the coal, rock from the roof and floor of the mine, and pyrite, an iron and sulfur compound. Transporting and handling these inorganic minerals along with the organic, heat-giving coal adds unnecessary expense, and their interaction with the power generation system imposes further problems. During combustion, for example, minerals create extra ash in the boiler. Mineral matter converts to slag, which coats boiler surfaces and acts as insulation, inhibiting proper transfer of heat to the water in the boiler tubes. Sulfur in the coal burns during combustion and forms SO_2 and SO_3.

Mineral matter in coal exacerbates costs of maintenance and outage not only in the boiler system but also in the emissions control systems. Ash, besides corroding, erodes fans, ducts, and tubes—having an effect like a light sandblasting. Last, but not least, of the problems associated with the inorganic content of fuel is the disposal of ash, which may take up valuable land space.

The cumulative costs of transporting, burning, removing, and disposing of stowaway impurities in coal continue to bloat the overall costs of electricity generation. Because impurities promote wear and maintenance problems, they reduce total annual operating time for a power plant and can take a slice off maximum capacity, both of which translate into economic losses. Stacked on top of these costs are possible expenditures for the purchase of higher-cost electricity from other sources during plant downtime, and even the need to build new capacity as older units wear out or are derated because of deterioration in coal quality.

Nicholas Esposito, now a project manager at Jersey Central Power & Light Co., was formerly a site manager at the Homer City plant of Pennsylvania Electric Co. A coal-cleaning plant was built to comply with EPA sulfur emission regulations, but Esposito, in retrospect, sees that another major benefit accompanied the change.

Two of the three Homer City units had design loads of 650 MW but had been restricted to 550 MW because of high ash levels and boiler slagging. When the cleaning plant reduced ash levels in the coal, these two units regained 30 MW each to achieve a level of 580 MW. In addition to moving toward compliance with the emission regulations, according to Esposito, the company effectively saved $60 million that might have been necessary to build 60 MW of generation replacement. Esposito adds, "EPRI sponsored some studies that helped us resolve some major problems in redesigning the coal-cleaning plant at Homer City."

Apart from the savings in operating costs that accrue from coal cleaning, a net increase in Btu/t of coal will also be possible through better cleaning. In conventional cleaning plants, 2% of the Btu are rejected with coarse coal and 3% with fines. In effect, 5 t out of every 100-t load are lost. Taking an average price of coal as $35/t, the real price rises to $36.75/t. Added to that is the cleaning cost itself, $2–$6/t of clean coal. Nevertheless, it has frequently been economical in the past to dispose of the fines because their collectively large surface area causes them to hold a high percentage of moisture. Moisture in coal not only creates weight and makes transportation more costly, but when coal is burned, it causes combustion heat to be dissipated through evaporation. In any coal supply, 15–25% could be fines, so the need for both effective separation from mineral matter and better dewatering methods is evident.

The test facility

Unlike commercial cleaning plants, EPRI's test facility is not dedicated to a single flowsheet or sequence of cleaning processes but is built to accommodate five basic flowsheets. In fact, with the addition, subtraction, or change in sequence of individual processing units, up to 50 different flowsheets can be set up to simulate a particular utility cleaning plant.

Each processing unit, such as a cyclone or filter, is flexible in that it can be adapted to process different quantities and sizes of coal. This means that fewer test units are required in the facility to simulate a wide range of cleaning-plant designs, thereby keeping testing costs down.

The processing units at the facility are of commercial size, so test results from the facility will be directly applicable to utility-scale cleaning plants.

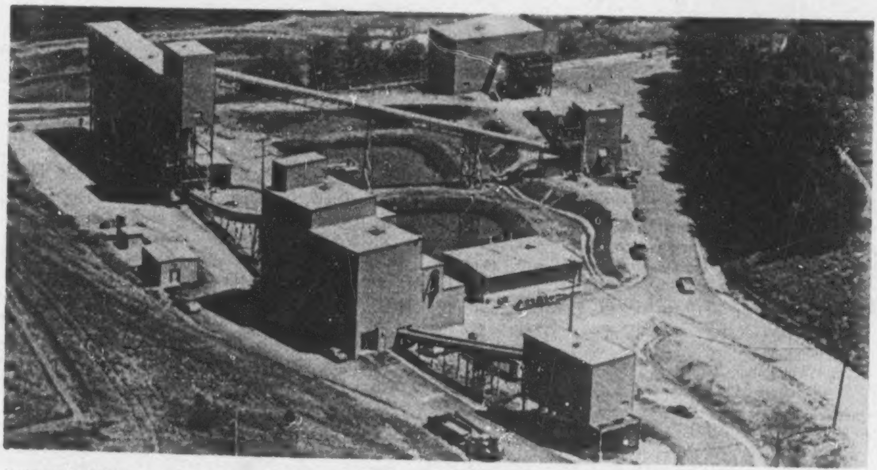
Electric utilities are invited by EPRI to participate in the test program by contributing run-of-mine coal samples of about 1000 t to the facility. Coal can be experimentally cleaned in up to five process sequences by running a fraction, say, 200 t, through a particular flowsheet at a time.

Participants will receive a reliable estimate of their sample's cleaning characteristics and a direct comparison between run-of-mine and clean coal. This estimate will be more valid than one based on a small laboratory sample, which is the general practice.

As Douglas Trerice, manager of the EPRI facility, explains, "If a utility is interested in building a coal-cleaning plant or understanding the cleanability and cost of cleaning its coal, the lowest-risk path is to test its coal at the test facility. That way, the utility is assured that the large sample really represents its coal reserves and that the coal will be tested in commercially available processes, using equipment now on the market."

The clean-coal samples can be transported to test combustors elsewhere for determination of burning characteristics. After test cleaning and test burning, a utility should be able to judge the effect the clean coal will have on an existing boiler or how to design a new boiler to best burn the clean coal.

Robert Frank, project manager for coal cleaning and for performance and reliability at the Tennessee Valley Authority, plans to have a large coal sample delivered to the EPRI test facility soon. The coal, which is mined close to TVA's Paradise plant, has a high clay content that makes fines very sticky and more difficult than most coal to clean. Frank says that



Coal-burning utilities from all over the United States can send samples of their coal to EPRI's Coal Cleaning Test Facility near Homer City, Pennsylvania, for cleanability characterization at commercial scale. From these analytic data, utility engineers and operators learn the type of cleaning circuit best suited to the unique characteristics of their coal supply and obtain clean coal samples for combustion testing.

up to now the cleaning of fines in froth flotation cells has been more an art than a science. "We believe that the EPRI test facility can give us some guidance in solving problems in the fine-coal circuit for these western Kentucky coals. There's more than one unknown in the puzzle."

Besides providing data on the cleaning characteristics of coal and on the design of processing units and systems, the test facility will introduce more-accurate procedures for sampling coal. Other new instrumentation includes devices for measuring the flow and density of slurries and an on-line analyzer for determining ash and sulfur concentrations. A national coal-quality data base will also be established that could eventually hold unique facts, such as the potential for sulfur reduction in various steam coals of national importance.

Cleaning processes

All processes now available for the physical cleaning of fine coal are housed at the test facility. Because the emphasis is on practical application and improvement of existing technology, chemical cleaning processes, which are at the research stage, are not included.

When coal arrives at the test facility's receiving station (by truck and rail), the first of 54 samplers throughout the plant removes an amount up to 10,000 lb of coal for washability analysis. Then the coal, whether hard anthracite or softer bituminous or lignite, is crushed at rates up to 100 t/h to a top size between ¾ and ¼ in, is sampled again, and stored in one of five 100-t bins.

Coal can be stored according to specific seam source, mine, or size or as a blend. The blending is an automatic process in which a conveyor belt with a mechanical tripping device travels between bins, depositing specific amounts of coal as a mix. Weigh feeders measure and control the amount taken from the bins as feed to the test plant.

Once laboratory staff have tested the raw coal for washability by immersing it in fluids of different specific gravities and observing what percentages float or sink, the rest of the coal can enter the plant at up to 20 t/h to follow the most appropriate of the five flowsheets. These will include some combination of coal-and-mineral separation processes, for instance, heavy-media cycloning, water-only cycloning, froth flotation, and tabling. Supporting equipment consists of screens, a classifying cyclone, thickening cyclones, centrifuges, a disk filter, and a thickener. To change flowsheets, the sequence of operations can be altered by diverters installed between operating units.

Slurry or solids samplers are positioned on each unit to sample all inlet and outlet flows. These samplers can be activated simultaneously to take snapshot samples of all material in the plant at a single moment or they can be programmed individually. The emphasis is on advancing the science of accurate and representative sampling. Existing cleaning plants are "blind" to the amount and composition of the material going into and coming out of their systems, and as yet have no slurry samplers. This year at the facility, sampling data will form the basis for computer models to help utilities predict the efficacy of different cleaning procedures.

A coal sample that has been crushed to a top size of ¾ in is first sprayed with water and vibrated on a 0.5-mm-screen deck. On average, about 25% of the coal passes through the deck as fines, while 75% remains on the deck to be processed through the heavy-media cyclone, the water-only cyclones, or the concentrating table. The fines go to the water-only cyclones and the froth flotation cells.

The heavy-media cyclone, which separates coal and mineral matter larger than 0.5 mm, is a 14-in-diam cone-shaped container with a vortex pipe leading out of the top and an outlet pipe at the bottom. The mixture of coal and mineral matter is first added to a suspension of water and magnetite (iron oxide ground to a very fine size) and is then spun in the cyclone. The heavy magnetite medium increases the specific gravity, or density, of the fluid so that the less-dense coal floats and can be separated from the heavier mineral matter.

When the coal slurry is fed into the heavy-media cyclone and spun, the magnetite and mineral matter gravitate to the sides of the cyclone and slide to the bottom, where they exit. The coal whirls down at first, but as it is a lighter material, it is drawn upward by the vacuum of the vortex action of the cyclone to the outlet at the top.

The specific gravity of the suspension can be increased from water's nominal value of 1 to the desired cleaning level by adjusting the amount of magnetite. Since coal's specific gravity is between 1.2 and 1.5, the water and magnetite medium is usually kept between 1.3 and 1.8 to ensure that the coal will float; the mineral matter's average specific gravity is about 2.7, so it sinks in the magnetite medium.

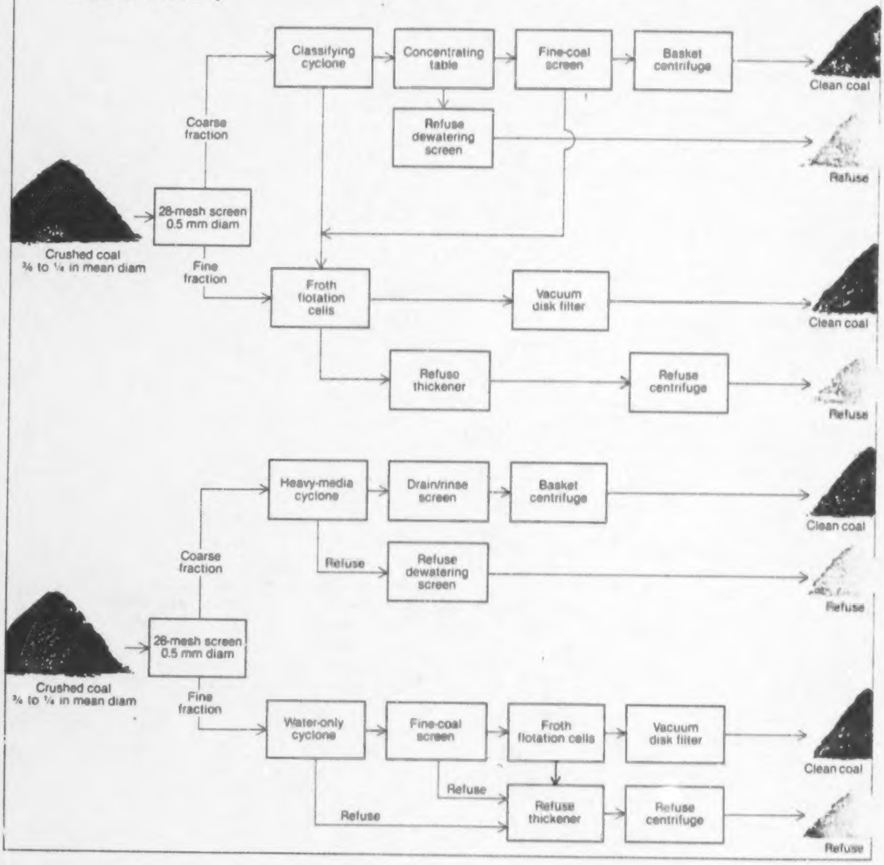
After leaving the cyclone, the magnetite is washed off the coal and refuse so that it can be reused. Refuse is dewatered (both disk filter and centrifuges can be used) and sent to disposal. The coal exits to a basket centrifuge for dewatering.

The concentrating table is designed to separate the larger-size fraction of coal and mineral matter by a method similar to panning for gold. It consists of a large, flat, tilted table with a grooved surface. The table is continuously vibrated while a stream of coal and water flows down its surface from the high side. The lighter particles of coal tend to concentrate on top of the mineral particles in the grooves and so are swept down the table to a collection area. The heavier mineral particles remain in the grooves and work their way to the side of the table, where they drop off to a refuse area.

The water-only cyclones are similar in shape to the heavy-media cyclone with a side inlet and outlets at top and bottom. The principle of separation is also the same—when the coal-and-water slurry enters the cyclone in a rapid flow, the heavy mineral particles gravitate to the bottom outlet, and lighter coal particles

THE COAL-CLEANING CIRCUITS

The three generic processes in coal cleaning are size classification, the cleaning process (separation from mineral refuse by a liquid medium), and dewatering/drying of clean coal for storage and of refuse for disposal. Five flowsheets are in place at EPRI's test facility, although numerous others can be simulated by simple piping changes. The two examples shown here represent only some of the circuits available. In the first flowsheet, coarser coal is cleaned on a concentrating table; finer coal goes to a froth flotation cell for separation from fine refuse in a method based on differences in the surface properties of coal and refuse. In the second flowsheet, the coarser fraction of coal is spun in a heavy-media cyclone, while the finer fraction goes through water-only cyclones and then proceeds through the froth flotation process. In both flowsheets, coarse coal is dewatered in a basket centrifuge and fines on a vacuum disk filter; coarse refuse is dewatered on screens and fine refuse in a centrifuge. Because the cleaning process is water-based, clean coal has a higher moisture content than raw coal. But clean coal compensates by containing less ash and sulfur per ton, which cuts down on wear of boilers, ductwork, fans, and flue gas emission controls and therefore increases plant availability.


Classifying cyclone
Concentrating table
Fine-coal screen
Basket centrifuge
Clean coal
Coarse fraction
Refuse dewatering screen
Refuse
28-mesh screen 0.5 mm diam
Crushed coal ¾ to ¼ in mean diam
Fine fraction
Froth flotation cells
Vacuum disk filter
Clean coal
Refuse thickener
Refuse centrifuge
Refuse
Heavy-media cyclone
Drain/rinse screen
Basket centrifuge
Clean coal
Coarse fraction
Refuse
Refuse dewatering screen
Refuse
28-mesh screen 0.5 mm diam
Crushed coal ¾ to ¼ in mean diam
Fine fraction
Water-only cyclone
Fine-coal screen
Froth flotation cells
Vacuum disk filter
Clean coal
Refuse
Refuse
Refuse thickener
Refuse centrifuge
Refuse

are flung through the top outlet. The water-only cyclones can be used for cleaning either coarse or fine coals in one or two cycloning stages. However, the finer the coal, the more difficult the cleaning process.

Froth flotation is the method most often used to separate very fine coal from minerals. The slurry is pumped into 20-ft^3 flotation cells after being agitated in a small storage tank where the percentage of solids in the slurry can be monitored for consistency to ensure the cell's efficiency is at its highest. Two banks of four flotation cells allow investigation of a wide range of flotation schemes.

The factor that differs from other coal-cleaning units is that oil is added to the slurry; because of its chemical similarity to coal, the oil adheres easily to the coal but not to the mineral matter. The oil allows air bubbles that have been injected into the flotation cell fluid to attach themselves preferentially to the coal particles. The fines then become so buoyant that they rise to the surface of the fluid and can be skimmed off the top, leaving the mineral particles behind. After the coal has gone through the fines cleaning, it is dewatered on vacuum filters.

Even with dewatering, clean coal ends up with a higher moisture content than raw coal. However, the trade-off is that it has a lower ash and sulfur content and more Btu per ton. This means that high-ash, high-sulfur Appalachian and U.S. interior coals become more attractive because flue gas emissions can be lessened with precombustion cleaning. The Clean Air Act Amendments of 1977 allow any combination of precombustion desulfurization of coal and postcombustion gas cleanup to meet SO_2 limits, so some utilities could find coal cleaning essential if they anticipate new local air emission regulations or contemplate the conversion of oil-fired plants to coal.

"The quality of the coal leaving any coal-cleaning plant depends not only on the coal itself but also on the flexibility of process control," says Trerice. EPRI's facility will allow development of completely automated coal-cleaning systems. All motors and instrumentation at the facility are connected to a digital computer that activates startup and shutdown and monitors the recording of operating data. This computerized system makes modifications in process control simpler than conventional systems. It eliminates the time lags typical in most cleaning plants between the time when corrective actions are needed and the time they are implemented with the computerized system, allowing an estimated 5% increase in recovery of Btu per ton of coal.

Comparative cleaning levels

Further savings have been demonstrated in a recent EPRI study of seven hypothetical coal-fired power plant systems, each using one coal from a particular mine and three levels of coal cleaning. Level A entails crushing and screening, with rejection of coarse mine debris. Level B represents the cleaning of coarse coal only (larger than ¼ in diam). Level C involves intensive cleaning—the coal is broken into two or three size fractions, each of which is cleaned separately.

Results of the study are encouraging. For each 1000 MW (e), the annual cost savings in coal transportation, handling, and waste disposal were in the range of \$1–\$7 million. The ash content in the coal shrank by between 10% and 70%, and 20–60% of the sulfur was removed.

The capital requirement for the Level B (partial) cleaning plants was about 2% of the total capital cost of the 1000-MW (e) coal-fired power plants (in mid-1978 \$). When using Level C (intensive cleaning), the share of power plant capital cost was 3–5%.

Frederick Karlson, manager of EPRI's Coal Quality Program, comments, "An interesting conclusion of the study is that in most cases the total capital investment is less when coal cleaning is used than when it is not. The capital investment required for coal cleaning can be more than offset by power plant savings. These savings result because clean coal generally requires a smaller boiler and reduced flue gas desulfurization."

The capital investment required even for a sophisticated cleaning plant is less than \$50/kW, whereas that required for a new power plant is more than \$800/kW.

The study showed that most new power plants could benefit from coal cleaning. This is particularly true if plant availability is improved. In fact, for the 1000-MW plants in the study, each 1% gain in availability translated into an estimated savings of at least \$1.5 million a year.

Growth in coal use

The Coal Cleaning Test Facility, as part of EPRI's Coal Quality Program, is becoming one of the major sources of national near-term R&D in this field, now that DOE's coal-cleaning program is being reduced and directed toward long-term R&D. In the event that the test facility is called upon to undertake even more R&D than was originally planned, it is well able to adapt to this kind of change because of the planned allowance for added systems and the flexibility in equipment combinations.

Considering that by 1995 new growth alone in utility coal consumption will be nearly three times greater than total coal consumption by the industrial sector, the opportunity to boost the heating value of this key fuel and to reduce the wear and tear in plant systems looks more worthwhile every year. Karlson estimates that by 1990 the electric utility industry will be spending about \$3 billion per month for coal. "The work of EPRI's Coal Cleaning Test Facility is significant not so much in cleaning coal per se," he says, "but for minimizing electricity generation costs." With the financial curbs now evident in the utility industry, coal cleaning can also go toward maximizing the availability of existing plants and so postpone capital spending for new units. ■

This article was written by Jenny Hopkinson. Technical background information was provided by Frederick Karlson and Douglas Trerice, Coal Combustion Systems Division.

Several processes of coal separation are discussed in this Noyes Data report, along with pictoral representations of these systems. Dewatering processes and water clarification are important considerations in coal washing, to achieve effective sulfur removal and reduce other forms of pollution. The great fluctuation in coal properties, such as sulfur and ash content, has created the need for several levels of coal preparation. The level of cleaning depends on the specific properties of the coal and what the coal will be used for.

Chapter 4

Methods of Coal Preparation

Coal preparation technologies have been used for over 100 years. Since the passage of the Clean Air Act in 1970, however, they have taken on greater significance as a limited means of removing sulfur from coal prior to combustion and reducing emissions of polluting sulfur dioxide. In general, coal preparation yields a more uniform quality fuel of greater heating value (Btu's) by reducing ash and trace elements (in some instances), removing mining wastes, and concentrating fixed carbon.

"Coal preparation" is frequently used loosely to describe all of the seven processes that coal undergoes between mining and combustion. These include transportation, storage, crushing, classification by size, separation of impurities, removal of water, and disposal of waste. However, for the purposes of this study, the definition of "coal preparation," also commonly termed "benefication," "cleaning" or "washing,"* will be narrowed to only those physical and chemical processes that upgrade the quality of the fuel by regulating its size and reducing the quantities of ash, sulfur and other impurities found in it.[1]

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*In the past, "cleaning" referred to the dry coal preparation technologies and "washing" referred to the wet coal preparation technologies. Today, the two terms are used interchangeably.

The chemical coal cleaning technologies promise greater total sulfur removal, although they are neither technically nor economically feasible at this time (for further discussion of the chemical coal cleaning process, see the Chemical Process chapter). Dry physical coal cleaning technologies, such as pneumatic cleaning and electrostatic separation, have also been used in the past without great commercial success.[2] Further, few of the dry cleaning technologies have demonstrated commercial potential, with the possible exception of the magnetic separation technologies (for further discussion of dry physical coal cleaning and magnetic separation technologies, see the Advanced Physical Coal Cleaning chapter).

At the present time, only the physical coal cleaning processes are used commercially.

Physical Coal Cleaning Processes

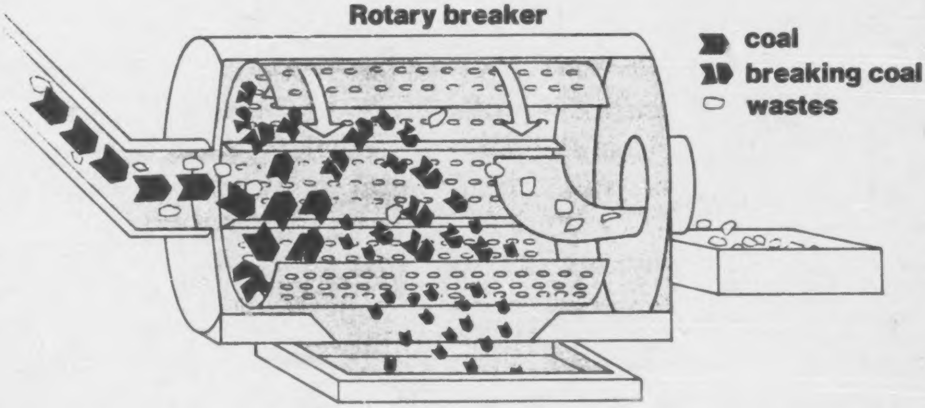
Today, coal preparation plants incorporate a variety of systems, ranging from those designed simply to remove coarse refuse from raw coal to sophisticated systems designed to remove the maximum amount of pyritic sulfur and ash. Thus a coal cleaning plant does not employ a specific process, but rather a number of different operations applied sequentially or in various combinations. A modern coal cleaning plant is a "continuum of technologies rather than one distinct technology."[3]

Coal to be cleaned is usually separated into three size fractions: coarse, intermediate and fine.[4] Preparing coals for more extensive cleaning involves crushing the raw coal feed to smaller size fractions and cleaning the coal particles of smaller and smaller size fractions.[5] Additionally, each fraction can be treated with more specific and sophisticated methods and equipment.

The six major operations are crushing, sizing, cleaning, dewatering, clarifying and drying.[6,7] The equipment available to accomplish these operations is produced by different manufacturers, and each unit has its own site specific applications. These different operations and variations are described below.

Crushing

Crushing, or comminution, liberates the larger impurities embedded in raw coal, and provides a product of more uniform size, facilitating the ease of transportation and proper combustion.[8] Over the years, there has been a great increase in the amount of coal which is crushed after mining.[9] In 1940, only 8 percent of the total U.S. bituminous coal production was mechanically crushed.[10] By 1975, this figure had risen to 90 percent.[11] Recently there has also been a general increase in the size of crushing and grinding equipment, which has made it possible for larger volumes of coal to be cleaned.



Sizing

Sizing separates feed coal into various fractions suited for washing by different cleaning processes. Sizing can be accomplished by wet or dry screening or classification equipment, which use variations in the flow rates of different size particles through a fluid or air. Sizing of particles above 28 mesh (½ mm) is generally performed on vibrating screens. For smaller particles of coal, sieve bends and classifying cyclones are used.[12,13,14]

Cleaning

The specific gravity of clean coal is less than that of ash, pyritic sulfur or the other impurities found in raw coal.* This difference in specific gravity is used by almost all of the commercially-used wet physical coal cleaning technologies to separate larger-sized coal particles (larger than ½ mm) from their associated impurities.

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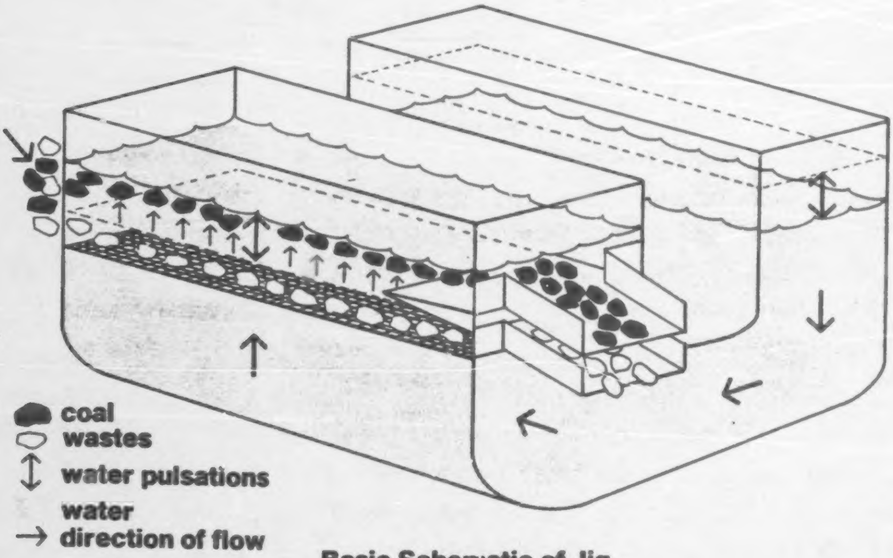
*Most coals have a specific gravity that ranges from 1.12 to 1.70. Ash and other impurities have a wide range of specific gravities: Pyritic sulfur—4.6 to 5.2; gypsum, kaolinite and calcite (typical ash constituents)—2.3, 2.6, and 2.7; and sandstone, clay and shale (typical mining wastes)—2.6.[15,16]

These specific gravity-based technologies, however, are usually not effective in cleaning raw fine-size coal particles smaller than ½ mm in size. When the size of coal is reduced, the surface area of the resulting fine coal particles is increased. Because of this increase in surface area, the separation times are longer for the fine particles. The pyrite in coal is often found in small, highly disseminated particles requiring extensive size reductions of the raw coal to facilitate its removal.

Seven processes are commonly used to clean coal:

Jigs are one of the oldest, and still the most common technology used for washing, separating and concentrating mineral ores. Introduced in Pennsylvania in 1873 for washing coal, jigs are used today to treat materials with a wide range of specific gravities, including gold (which has a specific gravity of 19.0). Jigging involves stratifying raw coal in water by pulsations that rearrange the particles so that the lightest (the coal) move to the top of the bed, while the heaviest particles (the impurities) settle to the bottom; the two products can then be collected separately.[17]

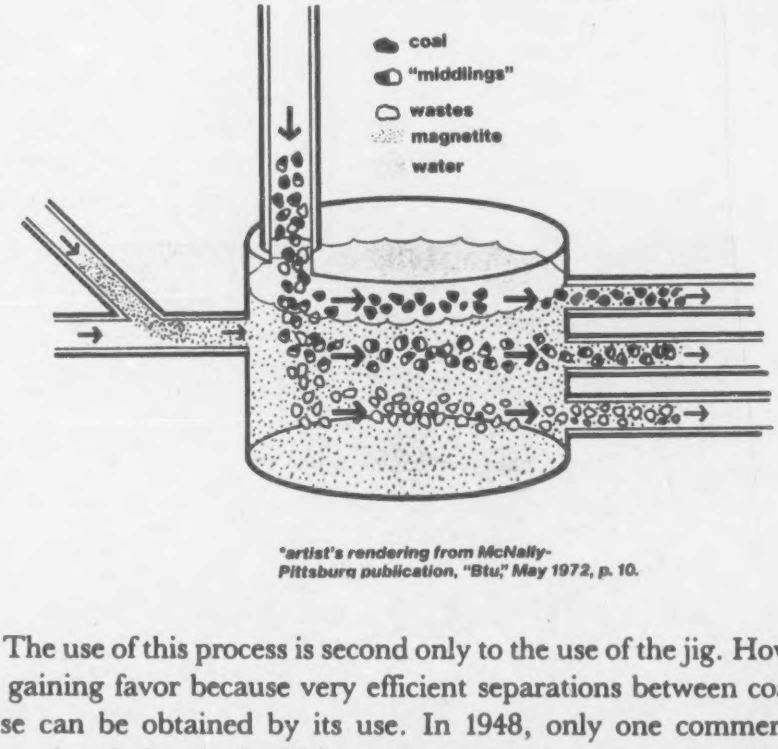


Basic Schematic of Jig

Innovations in design have kept jigs competitive with other cleaning devices. For example, the Batac jig, which was recently introduced in the United States, can process finer-sized coals more efficiently than the conventional Baum jig.[18]

Dense-media vessels use liquids that are heavier than water to separate coal from its impurities. Particles of coal as small as ¼ inch in size can be treated with this process. Finely ground magnetite (or other materials) mixed with water is used to make a slurry with the specific gravity required for the particular separation between coal and refuse. The specific gravity of the separating medium can be adjusted by varying the concentration of the magnetite to compensate for the variability in the specific gravities of the impurities contained within the coal, permitting the separation at the specific gravity needed to achieve the desired result.[19] The specific gravity can be varied to maintain a constant product, even with variation in the amount of impurities contained in the raw coal.

Principle of Dense Media Vessel*



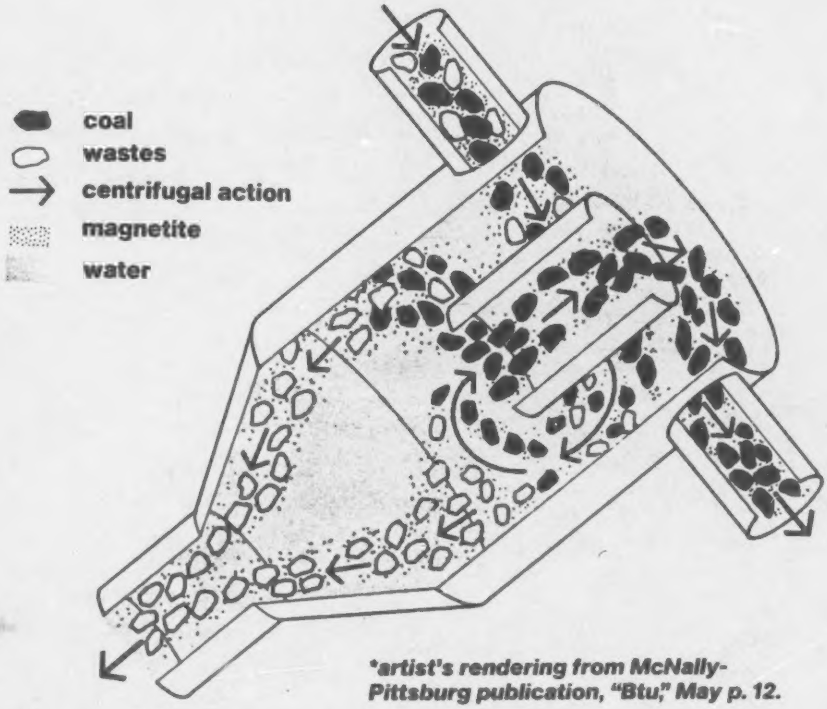
**artist's rendering from McNally-Pittsburg publication, "Btu," May 1972, p. 10.*

The use of this process is second only to the use of the jig. However, it is gaining favor because very efficient separations between coal and refuse can be obtained by its use. In 1948, only one commercially-operated coal plant in the U.S. used a heavy-medium process. By 1963, at least 115 plants in the United States were using this equipment to treat 49 million tons of coal annually.[20]

Magnetite recovery is an important part of this process, as nearly a pound of magnetite is used for each ton of raw coal. The amount of magnetite used depends largely on the size of the coal cleaned; the smaller the size of the particles to be cleaned, the greater the amount of magnetite consumed. The 1981 market value for magnetite was approximately 5¢ per pound.[21]

Dense-medium cyclones (revolving receptacles) are used to effect a sharper separation of intermediate sized raw coal material (½ inch x ½ mm). In this process, the gravity separation of a dense-media mixture (described above) is enhanced by centrifugal force.[22,23]

Dense Media Cyclone*



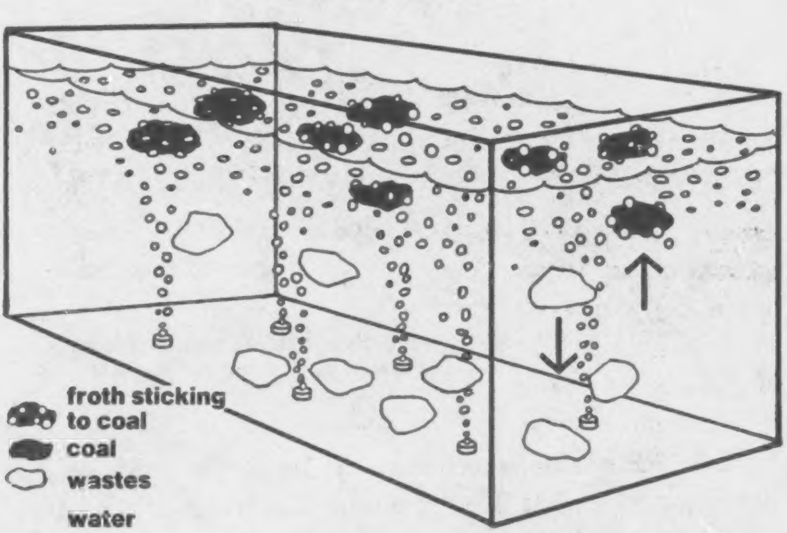
**artist's rendering from McNally-Pittsburg publication, "Btu," May p. 12.*

Hydrocyclones or water-only cyclones also enhance separation by centrifugal force, but they do not use a dense-media mixture. They are now being used more often in new preparation plants as rougher separating devices that decrease the load on downstream cleaning equipment. They are also used to clean very small flotation size coal (-28 mesh).[24,25] Hydrocyclone equipment is advantageous because it can process large tonnages of raw coal at a relatively low cost, and it requires comparatively less space and maintenance.[26] However, once installed, its power and water requirements are high.[27]

The concentrating table, a common hydraulic separator, is one of the oldest and most widely used cleaning devices in the United States. It is used especially for processing particles in the ⅜ inch to 0 size range. This method takes advantage of the fact that feed coal particles, "shaken" across a nearly horizontal surface, will separate from refuse because of their different densities.[28]

The froth flotation process is used to clean fine coal particles of less than ½ mm in size. Technologies used to remove the impurities associated with fine coal particles take advantage of the fact that coal, like oil, has surface properties that make it "hydrophobic" or water-hating. The most important of these technologies is froth flotation. In this process, raw fine coal particles are immersed in a water bath, through which air is bubbled. The water-hating coal particles cling to these air bubbles (nearly always with the help of flotation reagents) and can be skimmed off in a surface froth. The heavier coal impurities, which are not so hydrophobic, sink and mix with the water.[29]

Principle of Froth Flotation*



artist's rendering from McNally-Pittsburg publication, "Btu," May, 1972, p. 14.

The oil agglomeration process, which is still in the research and development stage, utilizes the same water-hating properties of coal particles, but uses oil droplets instead of flotation reagents. (This technology will be discussed in greater detail in the Ames Laboratory/Iowa State University profile in the advanced physical coal cleaning section.

Pneumatic cleaning processes such as air tables currently have limited application. Use of pneumatic cleaning equipment and methods decreased from about 14 percent in 1940 to about 2 percent in 1977 as a result of problems associated with dust, plant emissions, the higher moisture content of today's raw coals and the size limitations of pneumatic cleaning equipment.[30] But these methods of coal cleaning are attracting new interest in the west, where water is short and the coal contains fewer impurities. However, a large amount of research and development is needed to improve the efficiencies of these processes, and commercialization is not economically feasible at this time.

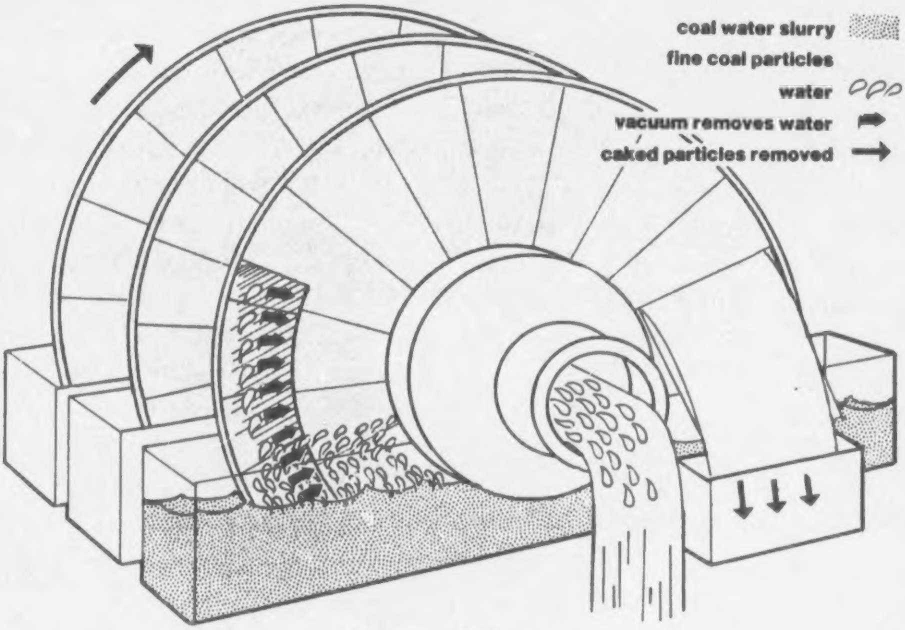
Dewatering

Wet physical coal cleaning processes leave a significant amount of moisture on the coal. This additional moisture decreases the fuel value of the coal, increases transportation and handling charges by adding to the weight of the coal shipment, and can make the coal prone to freezing in winter, causing problems in handling and use. To reduce the water content of the coal, preparation plants (in many cases) first dewater the coal mechanically, and then dry it by direct heating.

Screens are the most common type of fine coal dewatering equipment, and improved design in recent years has enhanced their dewatering capabilities. The "sieve bend" is now used throughout the world for preliminary dewatering before the coal passes to the vibrating screens and centrifuges. Another screen gaining popularity as a dewatering device is the "Vor-Siv," developed by Polish Coal, Inc. It is a stationary device, capable of handling high volumes of solids in a water slurry.[31]

Centrifugal dryers are used to dewater coal particles 1½ inch and smaller by spinning in a centrifuge. The effectiveness of this equipment depends upon the centrifugal force developed, the size of the particles, and the length of time the coal is left in the dryer.[32,33,34]

The vacuum filter is the most common dewatering device for fine-sized coal particles. Ninety percent of these filters are of the continuous vacuum type, which work well, are less expensive to purchase and install per square foot of filtration area and require less space than other filters. Solids form a cake on the leaves of this filter and are removed by blasts of air.[35]



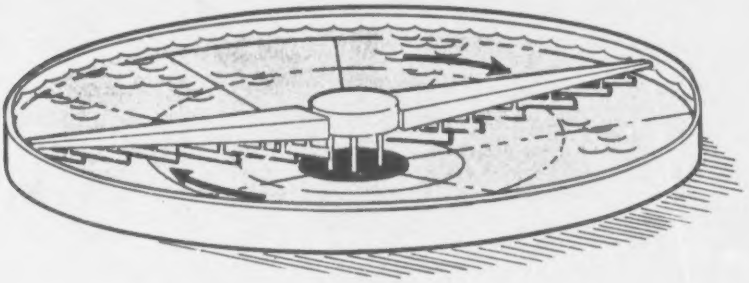
Disc Type Filter

Clarifying

Water Clarifiers are used in most preparation plants to remove nearly all of the solid impurities from process water. Preparation plants have very high recirculating requirements of up to 16,000 gallons per minute for a plant cleaning 1,200 tons of coal per hour (e.g. the Homer City plant). The total water requirements for all U.S. preparation plants is hundreds of billions of gallons per year.[36] Eighty percent of this water is recycled, and overall water management is of key importance.[37]

Currently, all wet coal preparation plants have some type of partially closed water circuit. Recirculation and clarification of wash waters is a critical part of cleaning plant operation because such treated waters decrease the loss of fine coal in the waste, prevent the build up of solids in the water recirculated to the cleaning units, and help the plant comply with state and local laws regarding stream pollution—which have become increasingly stringent in the last several years (see Appendix: Coal Preparation Plant Waste Disposal for a discussion of regulations governing preparation plant effluent streams). In the past, some preparation plants bled off a large portion of the polluting water into streams to keep the level of solids within the system low.

The gravitational clarifier (or thickener/clarifier) is the most commonly used piece of equipment for water clarification. It is a large circular basin in which solids settle to the bottom where they can be pumped away, leaving the clarified water to be removed at the top. The nature of the waste slimes and the need for clear water require that chemicals for flocculation be added to the water in the clarifier. This results in the formation of "flocs" (or clumps of wastes) and an increase in the settling rates of solids. Unless large amounts of clay are present in the wastes, requiring greater quantities of flocculates, the costs of these chemicals are generally more than adequately paid back by the reduced size of clarifiers and the enhanced clarity of plant water.[38,39]



Static thickener

Cyclones are used for clarifying plant water by controlling the solids concentration of the circulating water. A mixture of fine coal and water is concentrated by the centrifugal force of the cyclone to produce a thick suspension, which undergoes additional dewatering in filters or centrifuges.[40]

Thermal Drying

Thermal drying equipment removes moisture by passing hot gases over coal. It can absorb the surface moisture from cleaned coal, although the internal moisture is largely unaffected. Prior to 1967, the use of thermal dryers was increasing.[41] Since then, however, the number of dryers employed in preparation plants decreased as a result of stricter dust emission regulations which significantly increased the cost of the process. Only 11.8 percent of the mechanically-cleaned coal in the United States was thermally dried in 1977.[42]

Seven different types of thermal dryers are currently in use: the fluidized-bed,* multilouver, rotary dryer, screen dryer, suspension or flash dryer, and the vertical tray and cascade dryer.[43] The fluidized-bed dryer, introduced to the coal industry in 1955, represented more than half of the drying units in operation in 1981, processing over 72 percent of the coal to be dried.[44]

Levels of Coal Preparation

Not all coals are good candidates for coal cleaning, owing to differences in their constituents and physical properties. Also all coal preparation technologies have several problems and limitations that may reduce potential applications.

Depending upon the physical composition of the raw coal and the market requirements for cleaned coal, individual preparation plants may employ different combinations of the methods and equipment described above. The more fine coal particles cleaned, and the more sulfur, ash and other impurities removed, the more rigorous and expensive the levels of coal cleaning.[45]

The extent to which coal is cleaned divides all the coal preparation processes into different levels. There are many different classification schemes defining these possible levels. This report divides coal cleaning technologies, but they are actually only points along a continuum of possible levels of coal cleaning. In the real world of coal preparation, the distinctions between "levels" of coal preparation are often blurred.

The following table summarizes the five levels of coal preparation identified by INFORM.†

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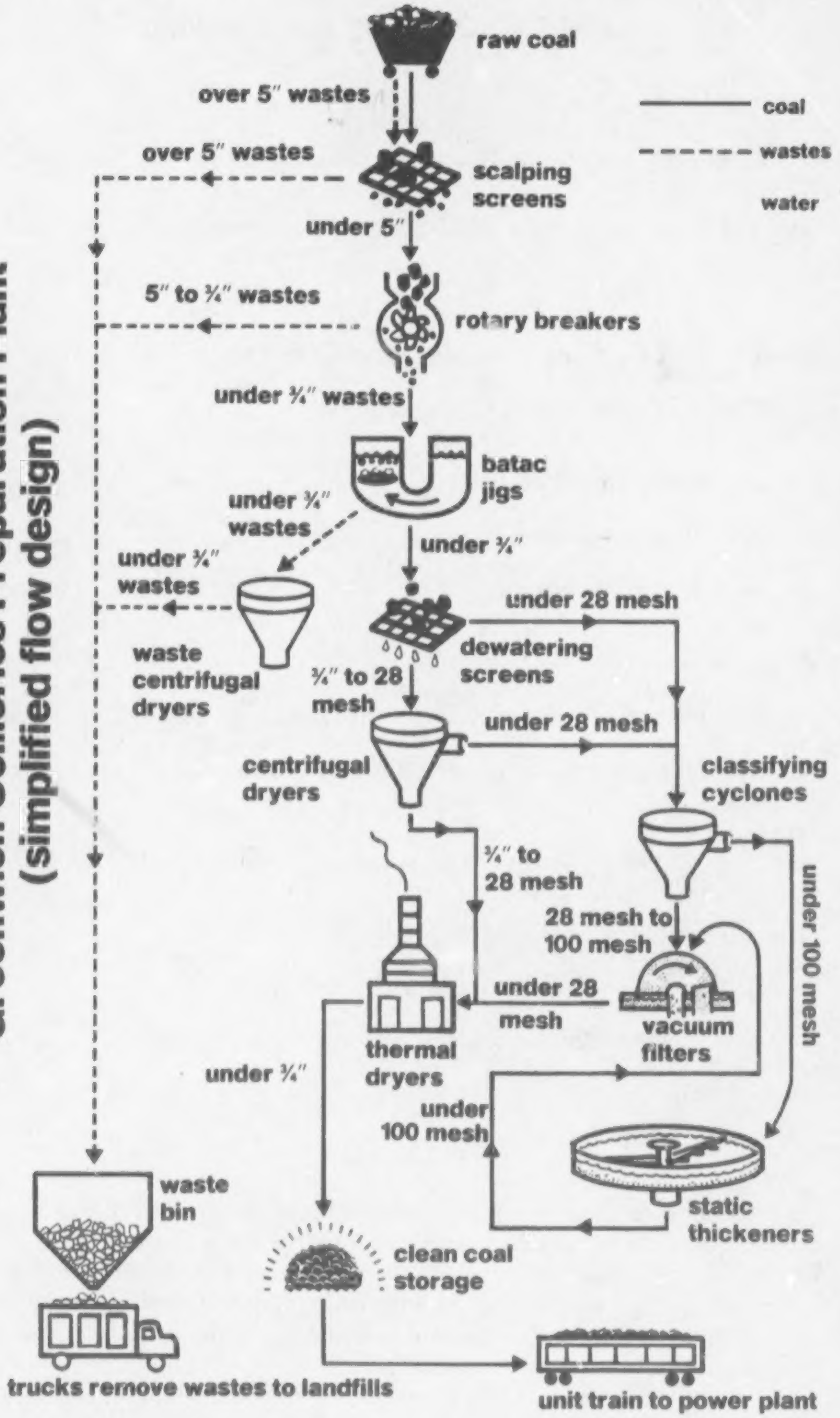
*See INFORM's study, *Fluidized-Bed Energy Technology: Coming to a Boil*

†The five levels described in this section are based on categories developed by Gibbs and Hill, Inc. in *Coal Preparation for Combustion and Conversion* (Palo Alto, California: Electric Power Research Institute, May 1978), pp. 2-49 through 2-64.


raw coal
over 5" wastes
coal
wastes
water
over 5" wastes
scalping screens
under 5"
5" to ¾" wastes
rotary breakers
under ¾" wastes
batac jigs
under ¾" wastes
under ¾"
under ¾" wastes
under 28 mesh
dewatering screens
waste centrifugal dryers
¾" to 28 mesh
under 28 mesh
centrifugal dryers
classifying cyclones
¾" to 28 mesh
under 100 mesh
28 mesh to 100 mesh
under 28 mesh
vacuum filters
thermal dryers
under ¾"
under 100 mesh
static thickeners
waste bin
clean coal storage
trucks remove wastes to landfills
unit train to power plant

Greenwich Collieries Preparation Plant
(simplified flow design)

Level 1 — Crushing and Breaking

Process Description:

Coal crushed to less than three inches (-3") for top size control.

Typical Percent of Raw Coal Weight Recovered:

98% to 100%

Percent of Raw Coal Btu Content Recovered:

99% to 100%

Typical Reduction Potential:

Ash: None to minor

Sulfur: Almost none

Benefits:

Removes very large-sized impurities such as wood, large rocks.

Smaller-size coal is better suited to transportation and handling needs.

Btu content of cleaned coal is almost 100% that of raw coal†.

Liabilities:

Ash and sulfur removal almost nonexistent.

Little increase in the heating value of the cleaned coal.

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†When coal is cleaned, some of the Btu content or heating value of the raw coal is lost in the refuse that is removed from the cleaned coal. However, the Btu value of the raw coal per pound is usually *less* than the Btu value of the cleaned coal per pound; this is due to the concentration of combustible materials in clean coal by the removal of ash and rocks that do not contribute to the heating value of the fuel. Thus, the Btu value per pound of cleaned coal may be higher than that of raw coal, although a portion of the *total* Btu content of the raw coal is lost in the discarded waste materials.

Level 2 — Coarse Coal Cleaning

Process Description:

Coal crushed to -3″.

3″ x ⅜″ "coarse" coal cleaned in equipment based on principles of specific gravity.

-⅜″ coal particles removed from raw coal and re-combined with 3″ x ⅜″* cleaned coal prior to shipping.

Percent of Raw Coal Weight Recovered:

75% to 85%

Percent of Raw Coal Btu Content Recovered:

90% to 95%

Typical Reduction Potential

Ash: Fair to good

Sulfur: None to minor

Benefits:

3″ x ⅜″ mining wastes (rocks) and some ash removed.

Heating value of the cleaned coal raised slightly.

Liabilities:

Little, if any, inorganic sulfur removed; coal not cleaned in small enough size-fractions to enable release of pyritic sulfur, except for large-sized pyritic sulfur impurities.

-⅜″ coal particles remain uncleaned and shipped with final cleaned coal product.

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*The sizes of coal particles (size-fractions) given in these tables are representative of those commonly found within the industry where size-fractions vary in different plants. Here, the term 3″ x 3/8″ refers to the coal that is between 3 inches and 3/8 inch in size. This abbreviation is used throughout these tables.

Level 3 — Moderate Coal Cleaning

Process Description

Coal crushed to -3".

Separated into three size fractions: 3" x ¼", ¼" x 28 mesh* and -28 mesh.

3" x ¼" coal cleaned in equipment based on principles of specific gravity.

¼" x 28 mesh particles receive more elaborate cleaning in specific-gravity equipment.

-28 mesh particles can be added uncleaned to other, cleaned fractions or discarded if they contain high percentages of sulfur and ash.

Percent of Raw Coal Weight Recovered

60% to 80%

Percent of Raw Coal Btu Content Recovered

80% to 90%†

Typical Reduction Potential

Ash: Good

Sulfur: Fair

---

*Mesh is a measure of size often used to describe small coal particles. A 28 mesh sieve has 28 holes or wires per linear inch, each opening equivalent to .595 mm. or .0234 inch. "-28 mesh" refers to all coal particles that are smaller than 28 mesh. When smaller sized coal particles with increased surface areas are washed, the moisture content of the cleaned coal increases geometrically. This increase in moisture may require thermal drying, which is energy intensive and therefore costly.

†More of the raw coal Btu content can be recovered if the -28 mesh fine particles are added to the final product.

Benefits

Cleaning of small size fractions accomplishes some inorganic sulfur removal.

Significant amount of ash and mining wastes removed.

Heating value of the cleaned coal raised significantly.

Liabilities

If -28 mesh uncleaned particles are added to cleaned coal, some sulfur and ash is reintroduced to the final product.

If -28 mesh uncleaned coal particles are not added to the final product, the Btu losses are high.

Level 4 — Fine Coal Cleaning

Process Description

Same as Level 3, except that -28 mesh particles are cleaned instead or discarded or added uncleaned to other cleaned fractions.*

-28 mesh particles cleaned either by specific gravity or surface property-based processes.

Percent of Raw Coal Weight Recovered

60% to 80%

Percent of Raw Coal Btu Content Recovered

80% to 90%

Typical Reduction Potential

Ash: Good to Excellent

Sulfur: Fair to Good

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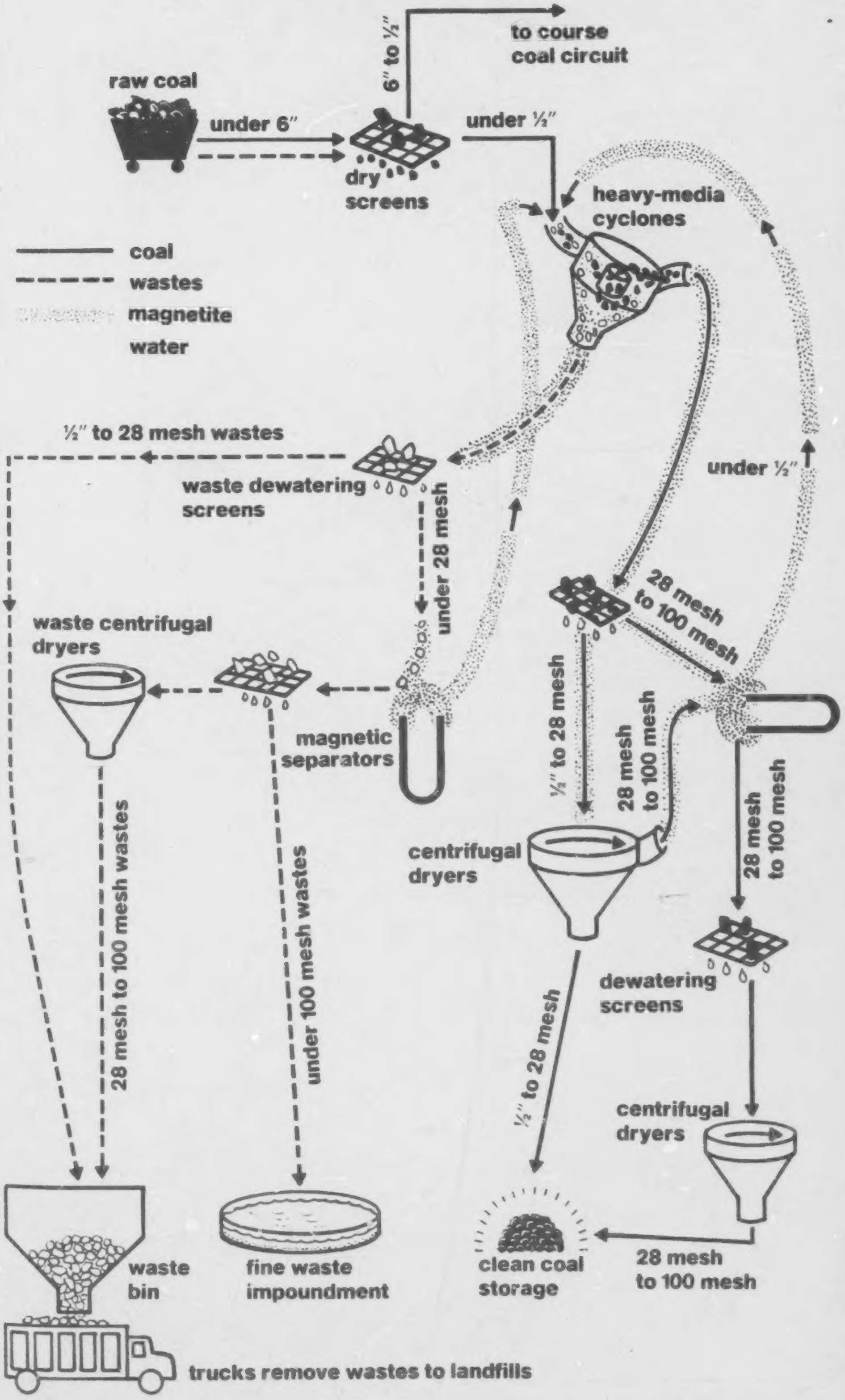
*In some Level 4 plants, raw coal is crushed to -3/4" (instead of -3") in order to release greater quantities of ash and organic sulfur from the raw coal.

Hamilton Preparation Plant—Fine Coal Circuit (simplified flow diagram)


raw coal
under 6″
6″ to ½″
to course coal circuit
dry screens
under ½″
heavy-media cyclones
coal
wastes
magnetite
water
½″ to 28 mesh wastes
waste dewatering screens
under 28 mesh
under ½″
28 mesh to 100 mesh
waste centrifugal dryers
magnetic separators
½″ to 28 mesh
28 mesh to 100 mesh
28 mesh to 100 mesh
centrifugal dryers
28 mesh to 100 mesh wastes
under 100 mesh wastes
dewatering screens
½″ to 28 mesh
centrifugal dryers
waste bin
fine waste impoundment
clean coal storage
28 mesh to 100 mesh
trucks remove wastes to landfills

Benefits

Significant percentage of inorganic sulfur removed.

Most of the ash (i.e., extraneous ash) is removed.

Heating value of the cleaned coal significantly raised.

Liabilities

Cleaning of smaller coal sizes results in increased costs (equipment, and operating and maintenance costs) and increased quantities of wastes that require disposal.

Moisture-related problems, as with Level 3; thermal drying may be required.

Btu's are lost in waste materials.

Level 5 — Multiple Stream Coal Cleaning

Process Description

Raw coal crushed and cleaned in 3" x ¼", ¼" x 28 mesh, and 28 mesh x 0.

Two distinct grades of cleaned coal produced: "deep cleaned," which is extremely low in sulfur and ash coal, and "middlings," higher sulfur and ash coal.

Low specific gravities used in "deep-cleaning" equipment to clean fine and ultrafine coal particles.

Percent of Raw Coal Weight Recovered

60% to 80%

Percent of Raw Coal Btu Content Recovered

85% to 95%

Typical Reduction Potential

Ash: Deep-cleaned coal—excellent; middlings coal—none to fair.

Sulfur: Deep-cleaned coal—excellent; middlings coal—none to fair.

Benefits

In some instances, deep-cleaned coal can be burned in plants that must meet stringent sulfur dioxide emission standards; middlings coal can be burned in plants that do not require extremely low sulfur or ash content in the fuel. The inorganic sulfur and ash content of most deep-cleaned coal is very low.

Fewer Btu's lost in wastes with Level 5 cleaning, as compared with levels 3 and 4; fewer combustible materials end up in wastes because the middlings, a higher sulfur and ash coal, is also recovered (an important consideration with costly coal).

Liabilities

High capital and operating and maintenance costs, relative to other levels of coal cleaning.

Problems related to use of a new technology.

Fine-coal related problems, as with Level 4.

Table A - LEVELS OF COAL PREPARATION

Level	*Raw Coal Weight(%)*	*Raw Coal Btu Content(%)*	*Reduction Potential*		*Comments*
			Ash	*Sulfur*	
1	98 to 100	99 to 100	None to minor	None	Crushing and breaking of raw coal to 3" size
2	75 to 85	90 to 95	Fair to good	None to minor	Coarse coal cleaning of 3" X 3/8" coal
3	60 to 80	80 to 90	Good	Fair	Moderate coal cleaning of 3" X 28 mesh coal
4	60 to 80	80 to 90	Good to excellent	Fair to good	Fine coal cleaning of 3" X 0 coal
5	60 to 80	85 to 95	Deep-cleaned coal: excellent; middle-cleaned coal: none to fair		Multiple-stream coal preparation of two cleaned coal products: "deep-cleaned" coal and "middle-grade" coal

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The goal of advanced physical coal cleaning is higher sulfur and ash removal at lower than current costs. These processes, several of which are listed and discussed on pages 173-174 of this Noyes report, are still in the developmental and pilot-plant stages.

Chapter 5

Advanced Physical Coal Cleaning

Advanced physical coal cleaning technologies are particularly innovative variations of the "conventional" physical coal cleaning processes that remove impurities from raw coal by crushing the fuel and treating it in a water slurry with jigs, cyclones, froth-flotation cells, and other equipment.

Advanced physical coal cleaning promises several advantages over conventional technologies: (1) higher percentages of ash and inorganic (or pyritic) sulfur can be removed; (2) hard-to-clean fine coal particles can be treated more effectively; (3) costly and energy-intensive "dewatering" and drying circuits can be avoided; and (4) the overall cost of cleaning may be reduced once the technologies are applied in full-scale commercial production.

However, several disadvantages may limit the usefulness of these techniques: (1) they cannot remove the organic sulfur from raw coal, as chemical coal cleaning attempts to do (see the chapter Chemical Processes); (2) the initial capital and operating and maintenance cost projections may not indicate the final cost of cleaning coal in full-scale operations; and (3) some processes may present problems of waste disposal and polluting emissions.

Some of the advanced physical coal cleaning technologies appear to be suited to commercial use. They are either in the experimental (laboratory) or pilot/demonstration-plant phase of development. Most

ADVANCED PHYSICAL COAL CLEANING TECHNOLOGIES

Developer *Process*	*Year Work Started*	*Sulfur Reduction Achieved* (Best test case)*
Advanced Energy Dynamics Electrostatic Separation	1979	37 to 68% total sulfur removal
Helix Corporation Clean Pellet Fuel	1975	2.12% sulfur raw coal; clean pellet emits 0.34 lbs sulfur per million Btu
Iowa State University Oil Agglomeration	1974	88% pyritic sulfur removal
Oak Ridge National Laboratory High Gradient Magnetic Separation Open Gradient Magnetic Separation	1976	60% pyritic sulfur removal-High Gradient Magnetic Separation (HGMS)
Otisca Industries, Ltd. Otisca Process	1972	70% pyritic sulfur removal (63% of the total sulfur removed)

Ash Reduction Achieved	*Technical Approach*	*Markets*	*Costs*
51 to 59% ash removal	Finely pulverized coal fed into modern iron ore cleaning modules that separate coal from wastes because of their differences in electrical conductivity	Utilities, industrial coal users	Company projects that electrostatic cleaning is 32 to 77% cheaper than "conventional" coal cleaning technologies
Not available	Raw coal combined with limestone to form burnable pellet. Pellet burned directly or gasified	Industrial boilers, furnaces in hospitals and prisons	Raw Coal: $1.20/million Btu; Clean Pellet Fuel: $2.00/million Btu
63% ash removal	Fine coal/water slurry combined with fuel oil to encourage separation of coal and wastes	Utilities, industrial coal users (that can burn a coal/oil mixture)	For theoretical 200 ton per hour plant, capital and fixed costs estimated at $1.50 per ton of raw coal processed (1974 dollars)
50-60% ash removal (HGMS)	Fine coal cleaned with dry, magnetic equipment that removes impurities from coal	Utilities, industrial coal users	For a 200 ton per hour HGMS plant: operating costs of $3.78 per ton of clean coal, with a capital investment of $23 million; comparable "conventional" plant would have operating costs of $4.97 per ton of cleaned coal and a capital investment of $16 million
87% ash removal	Heavy organic liquid (CCl_3F) used as cleaning bath instead of magnetite/water mixture to clean coal by gravity separation	Utilities, industrial coal users, metallurgical coal users, synfuel producers	For a projected 400 ton per hour plant: Capital costs of $6 million; Direct Operating Costs of $1.45 per ton of raw coal; Total Costs of $3.65 per ton of raw coal (1979 dollars)

of their developers were reluctant to predict exactly when they would be applied commercially. However, estimates of the period of time needed for commercialization range from one year or less to over five years. Of the technologies profiled, the Otisca Process is the closest to being commercialized—a demonstration plant is operating at the site of a utility-owned mine in Ohio. By contrast, the ''oil agglomeration'' process has so far only been tested under laboratory conditions at Iowa State University.

Five advanced physical coal cleaning technologies are profiled in this report: Clean Pellet Fuel, oil agglomeration, magnetic separation, the Otisca Process, and two miscellaneous techniques (electrostatic separation and the Kintyre Process). These technologies have been developed in several settings, including corporate, academic and government research laboratories. The sample represents some of the most sophisticated research in physical coal cleaning, and covers most of the novel technical approaches to cleaning coal by physical means. In three cases, other U.S. research groups have been investigating similar techniques. The groups chosen for this report either are further along than other researchers in developing the processes or offer greater promise in their technical achievements.

The preceding table summarizes the advanced physical technologies profiled.

The advanced physical coal cleaning technologies were developed during the early and middle 1970s. With the exception of Clean Pellet Fuel, these processes remove over half the ash and inorganic sulfur from the raw coal. Clean Pellet Fuel emits very little sulfur dioxide when burned, but leaves substantial quantities of ''bottom ash'' in the boilers. This fuel differs from the others (see Helix Technology Corporation profile). With the exception of the Kintyre Process, the remaining fuels are based on concepts that are closer to the conventional approaches to coal cleaning (see profiles of Other Technologies).

According to the organizations profiled, the markets for the advanced physical processes range from large utility boilers to small units located in hospitals, schools and prisons. Several technologies can provide fuel for users with special needs, such as the synfuels industry and utility or industrial boilers that burn a mixture of coal and oil.

Capital investment and operating cost estimates for the processes were provided by their developers. Although comparisons are difficult to make, the developers claim that their technologies can save more money than most conventional coal cleaning techniques. It is important to note that the advanced physical coal cleaning methods, like any new technology, are still untested in day-to-day operations in commercial settings, where unexpected difficulties may increase costs and decrease efficiency.

Chemical coal cleaning, when it becomes available, could eliminate the need for desulfurization systems at the user sites. Analysis of chemical cleaning, its advantages and disadvantages, and details of the process are presented. The table on pages 181-182 lists the various processes under development, along with the projected effective rates of SO2 and ash removal. This report was compiled by Noyes Data Corporation in 1981.

Chapter 6

Introduction to Chemical Coal Cleaning

On the forefront of the physical coal cleaning technologies today are innovations such as froth flotation units, multiple screening devices and magnetic separators; all of which use the difference between the physical properties of coal and its impurities to separate the two. Chemical coal cleaning (CCC) uses more sophisticated chemical processes to remove sulfur from coal, including the tiniest particles of pyritic sulfur and some of the organic sulfur which are beyond the removal ability of any physical coal cleaning process.

Present CCC technology can remove almost 100 percent of the pyritic sulfur from most coals. Total sulfur removal by various methods of CCC, including the removal of organic sulfur, now ranges from 50 to nearly 90 percent. Presently, CCC research and development is aimed at attaining the cost-effective removal of at least 90 percent of the sulfur in coal, thereby allowing new power plants to burn coal without scrubbers.[1]

According to organizations developing these processes, CCC is six to ten years away from commercial application.

In 1980, INFORM selected nine government and corporate organizations that were developing the most advanced chemical coal cleaning processes to remove pyritic and organic sulfur, and whose programs were expected to continue. Three of the programs studied, the Ames Laboratory, the Jet Propulsion Laboratory and the Pittsburgh Mining Technology Center, were government projects, and the rest were corporate. All of these programs, except ARCO, depend now, or did depend on government research funds from the Department of Energy (DOE).

Present Status of Research and Development

Most of the research into CCC began before 1977 when emission standards for utility boilers were less stringent than they are now. While government and corporate organizations have developed processes, designed plants and estimated their potential problems, benefits and costs, not a single CCC process has been developed to the full pilot plant stage; although eight of the developers claim to be ready for pilot plant programs which would cost from $5 to $10 million.

As a result of the 1977 amendments to the Clean Air Act and the new Administration's emphasis on private research and development, only two chemical coal cleaning methods, GE and TRW's gravimelt—the only processes that now promise to desulfurize coals to a degree which entirely eliminates the need for stack gas scrubbing at new utility plants—are now being funded by the DOE. Two other CCC projects studied, ARCO's and Kennecott's, are continuing with their own funds.[2]

The DOE budget for research into CCC decreased from $4.4 million in 1980 to $3.0 million in 1981, and further decreases are expected in 1982. The 1982 $1 million OMB appropriation for CCC has not yet been approved by both houses in Congress.[3]

Process Description

For the purposes of this discussion, chemical coal cleaning will include those methods of chemical coal cleaning that use regenerable industrial chemicals, in a procedure in which coal remains in the solid form throughout processing, and in which solid desulfurized coal is the only fuel product. Not included in this discussion are methods of coal treatment that partially or completely convert solid coal into another form. All of the methods of chemical coal cleaning described below have the potential to extract organic as well as pyritic sulfur.

In most chemical coal cleaning processes, coal is usually crushed to

-14 mesh.*[4] Chemical desulfurization techniques work best on coal which has also been physically cleaned (and crushed to fine or middling size) as large amounts of coarse pyrite actually inhibit the chemical reactions which remove organic sulfur.[5] Coal with pyrite and ash forming minerals also tend to use up more reagent in CCC than does cleaned coal, adding an unnecessary cost.[6] For these reasons, economic estimates of CCC assume that the coal feed is physically washed. This coal is mixed with water or a water and chemical reagent solution to form a slurry. The slurry and any additional chemical reagents are fed into a reactor vessel, which is sealed, heated and/or pressurized.† Under these conditions, much of the sulfur in the coal goes into solution with the liquid reagents.

All nine of the major chemical coal cleaning technologies presented in this report utilize either "oxidative" or "caustic" leaching to remove sulfur. In "oxidative" leaching, individual organic sulfur atoms are oxidized to form sulfur oxides (SO_x) or sulfates (SO_3), chemical groups that are easier to "cleave" or break from the rest of the coal molecular structure than unoxidized sulfur. "Caustic" leaching utilizes strongly basic substances such as lye, that actually displace organic sulfur atoms from the coal molecular matrix.

Dr. Robert Meyers of TRW has noted that a CCC reagent must have four specific attributes: 1) it must react very selectively with a coal's sulfur and not with the coal itself; 2) it must be highly recoverable from the coal after the desulfurization process; 3) it must be regenerable, so that most of it can be chemically processed back into its original form; and 4) it must be inexpensive, as a small amount will always be left on the coal and have to be replaced.[7] Satisfactory reagents are in use that remove pyritic sulfur and some organic sulfur.

Coal emerges from the reactor vessel impregnated with the sulfur-containing reagent fluid. It is washed with a water solution to remove the reagent fluid and is pressed into a "cake" and dried. The used slurry fluid is then passed over lime, which separates the water from the used reagent. This water is recycled back into the process and the used reagent is processed further to separate a sulfur containing waste from

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*This is much smaller than the size of particles treated in a conventional coal cleaning plant, where the smallest particles are generally 28 mesh.

†The GE process is an exception to this rule as it works on a dry coal, not on a slurry, and operates at ambient temperatures.

Company and Primary Business	*Name of Process*	*Research Began*
Ames Laboratory Government-funded research laboratory	Ames Oxydesulfurization (alkaline oxydesulfurization)	1974
Atlantic Richfield Company Diversified natural resource company	ARCO desulfurization	1976
Battelle Memorial Institute Diversified natural resource company	Battelle hydrothermal process	late 1960's
General Electric Electrical equipment, materials, natural resources, services and transportation equipment	GE microwave process	1975
Hydrocarbon Research Institute Research and engineering firm	Kennecott oxydesulfurization process	1979
Jet Propulsion Laboratory Government-sponsored space and energy laboratory	Chlorinolysis process JPL I uses chlorine gas JPL II uses water instead of chlorine	1975
Pittsburgh Energy Technology Center National energy research center	PETC oxydesulfurization process	1970
Research-Cottrell Energy and pollution control equipment	KVB - Guth process	1970
TRW, Inc. Research, Engineering and Manufacturing Firm	Gravimelt	1979

*Percentages are approximate and may vary within a few percentage points

Funds Available for Research	Sulfur Removed(%)			Ash Removed(%)	Btu Loss(%)
	Pyritic	Organic	Total		
$500,000 as of June 1980. Half from DOE and half from State of Iowa	95	30	70*	20	5
Estimated total exceeds $7 million. $1 million from EPRI, $6 million from ARCO	98*	35	70-80	NA	10
Several million dollars. Approximately 90% from Battelle and 10% from EPA	95	25-40	NA	NA	NA
Over $1 million. 80% by DOE, EPA, and NSF, and 20% by GE	Method 3 95*	60*	80*	NA	NA
Several million dollars invested by Kennecott. HRI also spent several man years in process engineering work. Recent government investment for testing by TRW (DOE)	Oxygen No. I 90	15-25	70	none	10
	Ammonia No.II 90	30-40	-	-	-
Approximately $1.4 million, 85% from DOE and remainder from Bureau of Mines and JPL			65* 70*	4	10
$1,285,000-- $400,000 in 1979 and $885,000 in 1980 from DOE	90*	40*	70	10	10
$665,000 - KVB provided $100,000 and Research Cottrell the remainder. May 1980, DOE grant of $415,000	95*	40*	50-70	NA	NA
$450,000--$307,000 from DOE contract awarded in May 1980	-	-	80*	50-95	5

as much of the reagent as possible. The purified reagent is then used again.

The reagent regeneration step produces chemical coal cleaning's principle by-product, gypsum. The wash water from all the processes profiled must be cleaned and reused to ensure economic viability, so the by-product gypsum is left relatively dry. Because the desulfurization step also removes pyrite (FeS_2), iron oxides are formed and end up in the gypsum as rust, making it brown in color. This gypsum also contains significant quantities of possibly toxic trace metals removed from the coal during desulfurization.

At present, it appears that techniques of caustic leaching have the potential to remove a higher percentage of sulfur than methods of oxidative leaching. Higher percentages of sulfur removal are also realized with both techniques when they are applied to coals with a lower organic and higher pyritic sulfur content.

This general approach to chemical desulfurization is the same for all of the processes profiled. Of course, variations on each step of the process outlined above occur within the individual techniques. For example, the GE process works on dry coal, not on a slurry; and the ARCO process uses a non-regenerable reagent.

Present Chemical Coal Cleaning Capabilities and Applications

As noted earlier, chemical coal cleaning can remove both the pyritic and organic sulfur compounds in coal, whereas physical coal cleaning can extract most of the pyritic sulfur, but none of the organic sulfur. The organic sulfur content of U.S. coals, however, does account for a substantial portion of the sulfur in coal:

Average Sulfur Content of 455 U.S. Coal Beds[8]
Representing Principal Coal Producing Regions

	Sulfur Content By Weight (%)
Total Sulfur:	3.02
Pyritic Sulfur (mineral):	1.91
Organic Sulfur (By Difference):	1.11

According to INFORM's sample, the most advanced physical coal-cleaning techniques remove at most 75 percent of this pyritic sulfur (or 47 percent of the total sulfur*). Thus, an "average" U.S. coal, physically cleaned by the best available practice, would still contain 1.4 percent total sulfur. (Some Appalachian and midwestern coals would nearly double this "average" figure.)

All of the CCC processes examined in this report have been proven to be highly effective in removing almost 100 percent of the finely dispersed pyrite left in coal after physical cleaning. This pyrite is not chemically bonded to coal molecules, but is present in such microscopically sized pores, that even in very finely crushed coal, water cannot wash it away. Chemical coal cleaning reagents can react with this pyrite, making it water soluble so that it can be easily removed. Chemical coal cleaning tests have shown removal capabilities from 50 to 88 percent of the total sulfur, including organic sulfur, in high sulfur coals. Three of the processes profiled, ARCO, GE and TRW's Gravimelt, have repeatedly achieved 80 percent total sulfur reductions in recent tests.[9] While these tests on high-sulfur eastern and midwestern coals are limited to bench scale experiments, they indicate the possibility of significantly desulfurizing U.S. coals.

Methods of CCC presently under development can also remove significant quantities of trace elements and some nitrogen in addition to the sulfur. This reduces the amount of possibly dangerous trace elements and nitrogen oxides emissions released during combustion. However, some chemical coal cleaning processes do not remove ash as effectively as does physical coal cleaning, and therefore CCC works best on coal that has been partially cleaned, with the coarse pyrite crushed and the coal reduced to fine and middling size.

Chemical coal cleaning plants could be built using existing flow sheets and almost entirely with proven industrial parts. The only new component that would have to be designed would be the desulfurization vessels. All of the other components, slurry pumps, coal pulverizers, driers, thickeners and filters, are proven technologies.

The environmental and economic importance of "clean" coal is a result of EPA's Revised New Sources Performance Standards (RNSPS) for utility boilers, which requires the removal of at least 90 percent of the sulfur present in a coal except where emissions are held to 0.6

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*The Homer City preparation plant began achieving this sulfur reduction in March of 1981.

pounds of sulfur dioxide or less; and then, 70 percent of the sulfur originally present in coal must be removed.

Data on two representative high sulfur coals provided by DOE illustrates the possible significance of CCC in meeting this EPA regulation:

COAL TYPE[10,11]

	Kentucky No. 11		*Illinois No. 6*	
	Total Sulfur (%)	*$SO_2/10^6$ Btu(lbs)*	*Total Sulfur (%)*	*$SO_2/10^6$ Btu(lbs)*
Raw Coal (typical) (1):	4.40	6.9	3.00	4.8
Physical Cleaning Only (2) (typical):	3.00	4.5	2.54	3.8
Sulfur Reduction(%):	31	-	15	-
Combined Physical and Chemical Cleaning (3) (experimental):	.53	0.80	.36	.56
Sulfur Reduction(%):	87	-	88	-

In terms of potential sulfur dioxide emissions and stack gas scrubbing necessary to comply with the New Source Performance Standards, the Illinois No. 6 coal conditions compare as follows:

Coal Condition	*$SO_2/10^6$ Btu Which Must Be Scrubbed*	*FGD Sludge Produced In[12] Scrubbing(Tons/Year)*
Raw:	3.38	140,000
Physically Cleaned:	3.24	132,000
Combined Physically-Chemically Cleaned:	None	None

These figures apply to a new 500 Mw plant, with a thermal efficiency of 33 percent and an 85 percent capacity factor.

In addition to providing a cleaned fuel for utility boilers, the chemical coal cleaning process developers also see small coal-fired boilers, generally termed industrial boilers, as a potential market. For such boilers, which need not meet the RNSPS, chemical coal cleaning could be a more economical method than scrubbers of meeting emission requirements. Industrial users would thereby avoid the necessity of finding the space required by scrubbers, and could avoid a scrubber's waste disposal problems. Furthermore, industrial boilers, unlike most utility boilers, are often turned off and on intermittently, which can cause problems in the use of an FGD system (see INFORM's *The Scrubber Strategy*), but would have no effect on the use of chemically cleaned coal.

In 1977, industrial boilers used approximately ten percent of all U.S. coal. A DOE spokesman estimates that growth in the industrial marketplace and conversions of industrial boilers to coal in the next decade could make the market for CCC as much as 200 million tons per year.

Problems of Chemical Coal Cleaning Technology

Chemical coal cleaning does create environmental hazards whose impacts must be minimized. Coal storage and handling presents a dust and particulate problem similar to that of physical cleaning but one that is aggravated by the high volume of very fine sizes handled in chemical coal cleaning plants.[13] Though sulfur is usually carried away from CCC reactors in liquid solution, some research has suggested that sulfur-containing gas is also produced by the desulfurization reactions.[14] This gas, probably mostly sulfur dioxide, although other gases are also present, would escape into the atmosphere when the sealed reactor is vented. In any commercial CCC plant, such gases might have to be scrubbed to meet EPA regulations—an ironic dilemma for a technology which was created as an alternative to FGD. Some developers claim desulfurization conditions can be adjusted to minimize these "off-gases."[15,16,17] However, more research is needed to fully assess the potential environmental impact.

Eight of the nine developers profiled anticipate marketing their "abatement" gypsum (ARCO is the exception).[18] However, the market possibilities for this product are unclear. The wallboard industry utilizes 73 percent of all U.S. gypsum, but the brown CCC product would not be an acceptable substitute for white and inexpensive natural gypsum. The cement industry also utilizes gypsum, but it is uncertain whether that industry could absorb the quantities of gypsum produced in large volume coal processing.

The alternative to marketing by-product gypsum is to dispose of it in a landfill where the leaching of trace metals would have to be controlled. However, the circumstances under which CCC waste leaches have yet to be established, and dangers of disposal will only become clear after complete pH and compositional analyses have been done on large quantities of actual refuse, i.e., after scaled-up research.

Cost Estimates

The primary consideration affecting chemical coal cleaning development planning is the cost effectiveness of the technology. Chemical coal cleaning will cost more than physical cleaning. Basic equipment costs will be higher and in addition, chemical reagents, recycling, and post-treatment requirements can be expected to add further expense.

To be cost effective, therefore, chemical cleaning must provide considerably greater benefit, in terms of the ultimate cost of energy, than physical cleaning. In other words, the ''cost'' of chemical cleaning cannot exceed its ''worth'' to the user.

Preliminary estimates of the costs of the nine methods studied range from $0.61/10^6$ Btu for scrubbers, and $3.00 to $4.00/10^6$ Btu for liquefaction and gasificaction.[19,20,21,22] Chemical coal cleaning is 10 to 30 percent more expensive than scrubbers and at least three to four times less expensive than coal liquefaction and gasification.

However, this economic comparison does not take into account the benefits that CCC could provide in addition to eliminating the need for flue gas scrubbing at a power plant. An unpublished DOE report calculated the potential worth of CCC as follows:

1. The elimination of flue gas scrubbing in a modern 500 Mw power plant could be worth as much as $30 per ton of coal;

2. Further benefits realized in transportation and boiler operations could be worth another $20.00 per ton;*
3. On this basis, the potential value of a chemically cleaned "supercoal," would be around $80 per ton (assuming input coal at $30 per ton).†[23]

According to these DOE figures, to be commercially viable, CCC must remove enough sulfur to eliminate scrubbers, at a cost of less than $50 per ton. Preliminary cost estimate by Bechtel, Versar and others vary widely, from as low as $15 to $20 per ton to as much as $50 to $60 per ton.[24] Thus, according to the DOE, these cost estimates "do not appear to present intrinsic processing cost barriers which would preclude commercialization."[25]

However, as the DOE report mentions, these cost estimates have been derived from extrapolations of laboratory-scale experiments and should not, therefore, be viewed as anything other than rough order-of-magnitude calculations. Dependable estimates will require experience with small-scale prototypic hardware and could be lower or higher than these present estimates.

The costs of CCC vary because the specific process steps which are responsible for most of the cost of CCC differ from process to process. However, some general observations can be made by dividing the costs of chemical cleaning into electricity, reagents, and lime.

The largest cost for most CCC processes is the electricity used to heat and/or pressurize the desulfurization reactors, to operate the slurry pumps (which move coal/water/reagent mixtures through the CCC system), and to dry the coal.[26,27] These electricity costs equal approximately 40 percent of the annual operating budget of every chemical coal cleaning process profiled.[28,29]

However, techniques to decrease electricity usage are being designed

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*The report notes, however, that accurate assessments of these benefits (as with physical coal cleaning) require detailed knowledge of specific site conditions, i.e., size, age, design of boilers, how often the plant is used, location of plant, and other local considerations.

†In 1981 dollars, a chemically cleaned "supercoal" would be worth about $3.10/10^6 Btu, on this basis which is equivalent to fuel oil at $17.86 a barrel—about half today's oil cost (roughly $6.00/10^6 Btu).

Estimates For A Theoretical System
Costs (1980)*

(8,000 ton per day plant)

Company	*Total Capital*	*Annual Operating*	*Size of Latest Test*
Ames Laboratory	$277,000,000	$21,240,000 ($1.48/MMBtu)	1 liter autoclave on 10-20 coals
Atlantic Richfield Company	$191,000,000+	$58,400,000+ ($0.92/MMBtu)	200 lb/hr pilot plant
Battelle Memorial	$168,100,000**	$74,800,000** ($1.31/MMBtu)	1/3 ton per day mini-plant
General Electric	(Method 3) $102,000,000**	$35,900,000** (+$0.61/MMBtu)	Batch tests of 100-200 grams/minute
Hydrocarbon Research Institute	No. 1 (oxygen) $255,000,000	$18,480,000 ($1.41/MMBtu)	Batch tests (no size given)
Jet Propulsion Laboratory	JPL I: $269,000,000 (chlorinolysis process) (uses water solvent)	$29,770,000 ($1.76/MMBtu)	2 kilograms/hour continuous mini pilot plant
Pittsburgh Energy Technology	$301,000,000	$18,840,000 (+$1.68/MMBtu)	Stirred tank reactor (no size given)
Research-Cottrell	$208,000,000	$32,110,000 ($1.22/MMBtu)	100 gram, 10 mesh coal samples in bench and continuous batch reactors
TRW, Inc.	$73,460,000§	$52,300,000§ ($/MMBtu unavailable)	1 to 2-gallon batch reactors

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*All estimates given, unless starred, are by Bechtel National, Inc. Bechtel's 1980 estimates include costs for physical coal cleaning prior to treatment

+Estimates by ARCO and EPRI. Costs exclude physical coal cleaning

**Estimates by Versar, Inc. Costs include physical coal cleaning

§Estimates by TRW. Costs exclude physical coal cleaning into process flow sheets. For example, a possible alternative to electrically heated reactors is to burn some of the cleaned coal. However, the cost of using as much as 7 or 8 percent of the product tonnage may not be less than the cost of the electricity. Centrifuges to partially dewater coal would use less electricity than thermal dryers, but centrifuges cannot always bring washed coal to a moisture level acceptable for firing in a pulverized-fuel boiler.

The most significant cost of CCC after electricity is the replacement cost of used reagents. This cost is specific to each process and to the kind of reagent used, and in some cases, it can exceed the cost of electricity. While recovery and regeneration levels may be improved, 100 percent reagent recovery will never be obtained in the processes now under study.[30] Reagent replacement cost is expected to be a major consideration in the JPL 1, Guth, Battelle, ammoniacal Kennecott, and Gravimelt processes.[31,32,33]

Included in this reagent category are process gases which are used as reagents. Oxidative processes which occur in oxygen are sometimes called oxydesulfurization techniques. Oxydesulfurization processes will require large oxygen manufacturing plants adjacent to CCC facilities.[34] Since this oxygen will be manufactured on site, the cost is not determined by commercial markets as are those of regularly purchased chemicals. Still, oxygen is a major reagent cost in the Kennecott, Ames, and ARCO processes.[35,36] In some cases, the presence of large quantites of oxygen presents a danger of combustion when used with a highly combustible fuel such as coal. However, oxygen plants have been operated safely at the site of other industrial facilities.

The cost of lime accounts for an average of 20 percent of the annual operating budget of each process.[37] This cost is expected to remain relatively stable.

The Future of Chemical Coal Cleaning Technology

Although chemical coal cleaning will benefit from further research in several areas, improvement in its ability to remove organic sulfur to consistently achieve a 90 percent sulfur reduction is essential if CCC is ever to become a commercial, viable technology.

Lack of sufficient theoretical knowledge about the molecular structure of coal is a key factor which now limits the capacity of CCC to remove organic sulfur. Organic sulfur appears in many different forms within coal's complex molecular structure, not all of which are known. The problem encountered in CCC is that the different forms of organic sulfur cannot be separated from coal under the same reaction conditions.[38]

Organic sulfur is present in coal in two principle forms, chain and ring sulfur. Some researchers suggest that it is almost impossible to break down ring sulfur, and that whatever sulfur is being removed by CCC is of the chain type.[39] There is growing evidence that roughly half of the organic sulfur in coal is of the ring type.[40] For these reasons it appears that there may be an upper limit to organic sulfur removal by both caustic and oxidative leaching. Coals with a high percentage of organic sulfur may not even be able to reach as much as 50 percent desulfurization.[41,42,43,44]

A related problem which troubled many CCC developers profiled in this report is the inadequacy of the standardized American Standard Testing Association (ASTM) procedure for measuring organic sulfur. The ASTM test is not a direct determination of organic sulfur content but a determination by difference (total sulfur minus pyritic sulfur) and it does not provide researchers with sufficiently accurate data. Consequently, several programs are now under way to develop a direct method for determining the amount of organic sulfur in coal.[45]

Problems also surround the drying and transportation of the CCC product, which is usually P.C. grind* or finer. Drying this product presents many of the same problems as drying fines from conventional preparation plants, and some research is now focused on performing CCC on a coarser grind of coal.[46] Although chemically cleaned coal would probably be pressed into cakes or pelletized, the difficulties of shipping large quantities of such coal are still unknown.

During fiscal year 1982 (which began October 1, 1981), DOE will be analyzing the sulfur removal capabilities and cost projections for the two CCC processes it is now sponsoring (GE and TRW's gravimelt) to determine which, if any, should be sponsored further. Bench scale tests are being continued to determine the relative capabilities of these processes. According to DOE, at the present time TRW's fused salt process appears to provide the most complete desulfurization. DOE anticipates that one or both of the candidate processes will be taken into the next phase, the objective of which will be to resolve critical process design uncertainties. Process steps which present the highest cost risks (e.g., filtering, reaction kinetics, material recovery, etc.) will be studied with respect to plant design, product quality, and material requirements.[47]

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*P.C. grind or "pulverized coal" is defined to be 70 to 80 percent 200 mesh.

When this work is completed, DOE states that the design of a prototype, small scale (5 tons per day) "process feasibility unit" (PFU) will be undertaken. The goal of this project, according to Agency officials, is to demonstrate the commercial viability of CCC in order to eliminate the need for flue gas scrubbing of selected high-sulfur U.S. coal in their respective utility markets.[48]

Physical coal washing is a relatively inexpensive way to reduce SO2 emissions. This Noyes Data article details the costs of building, operating and maintaining coal preparation plants.

Costs of Coal Cleaning

The cost of building, operating and maintaining a coal preparation plant varies, depending on the size of the plant, the amount and complexity of its equipment and the degree of automation. Inflation has driven up the cost of newer plants, and maintenance costs are often higher during the "start-up" phase of plants than they are after final operating adjustments are made and the personnel running the plants have become more familiar with them. Operation and maintenance (O&M) cost information is the most sensitive area of inquiry for preparation plant owners, and nine of these in INFORM's sample considered such information proprietary.

Cost information compiled by INFORM is presented in Table G. The figures, however, are not readily comparable. Dollar values have changed since many of the plants were built, and capital cost figures were not always furnished in the same year (for example, 1979 dollars). O&M costs are given in early estimates as the cost per ton of either raw or cleaned coal.

Nine of the 12 preparation plant owners supplied capital cost estimates for their plants. These ranged from $2.2 million (Northwest Coal/Western Slope) to $97.2 million (PENELEC/Homer City) in 1979 dollars. In most cases, the cost of building the plants was directly related to their size. For example, R&F's Warnock plant, with a level 4 cleaning ability using heavy-media equipment and processing 1,000 tons of coal per hour, was much more expensive ($22 million) than the

Table G: COSTS OF PREPARATION PLANTS

Owner/ Plant	*Capital Costs*	*Operation and Maintenance Costs*	*Financing*
AEP Helper	$10 million (1977 dollars)	NA	NA
AT Massey Marrowbone	NA	NA	NA
Duquesne Warwick	$15 million (1979 dollars)	NA	Sale of Duquesne Light Company stocks and bonds
Island Creek Hamilton	Course plant: NA Fine plant: $6 million (1979 dollars)	$1.00 to $1.50 per ton of raw coal processed	Internal cash flow from parent company (Occidental Petroleum); amortized over 20 years; investment tax credits given by government for plant construction
Monterey No. 2	$25 million (1977 dollars) includes front/back loading equipment	NA	NA
Northwest Coal Western Slope	$2.2 million (1979 dollars)	NA	Capital funding from parent company (Northwest Energy Corporation); investment tax credits given by government for plant construction
Peabody Randolph	NA	NA	NA
PENELEC Homer City	$97.2 million (as of 12/81)	$13 million per year (estimate)	Bond sales and reinvestment of company profits
PP&L/PMC Greenwich	$13.7 million (original costs) $19.9 million (estimated costs in 1979 @8% increase per year)	$3.6 million (first six months of 1980; yearly estimate: $5.4 million; $2.75 per ton of cleaned coal	Leases/loans; initial financing: $9 million loans, $4 million leases; 20-25 year payback period; floating rate of interest on loans
Pittston Rum Creek	NA	NA	NA
R&F Warnock	$22 million (1978 dollars) (+ $8 million for front/back loading equipment and cleaned coal storage facilities)	NA	Capital funding from parent company (Shell Oil Company)
TVA Paradise	$44 million - preparation plant ($68 million - conveyors and auxillary equipment) (1981 dollars)	$1.39 per ton of cleaned coal	Electric power sales revenue; bond issue

Northwest Coal's Western Slope plant ($2.2 million), which cleans 400 tons of coal per hour only to level 3, using a jig.

Two plant owners (PENELEC and PP&L/PMC) provided detailed cost breakdowns that illustrate the various subcategories of preparation plant costs (see table H).

The $97.2 million capital cost of the PENELEC/Homer City plant was unusually high, not only because it is a large and complex plant, but also because of the extensive modifications that have been necessary to correct design errors. The plant's capital costs were originally projected to be about $55 million. (See Table H and Homer City profile for more information.)

PP&L/PMC also broke down the capital cost figures for their Greenwich plant. This preparation plant was built in two phases, with a coarse coal cleaning plant completed in 1971 and a fine coal cleaning

Table H:

Capital Cost Breakdown for PP&L/PMC-Greenwich Plant*

	Original Cost	*Estimated Cost**
Buildings & Structures:		
(Coarse circuit) - Acquired in 1971:	$ 1,129,680	$ 2,090,959
(Fine circuit) - Acquired in 1976:	3,573,507	4,501,589
Total Cost:	$ 4,703,187	$ 6,592,548
Raw Bin & Screening Structure (1971):	$1,842,230	$3,409,839
Coal Refuse & Handling; Clean Coal Storage (1971):	$1,189,669	$2,201,994
Machinery, Labor, Materials (1976):	$5,797,000	$7,302,550
Raw Coal Storage System (1971):	$ 188,969	$ 349,264
Total Plant Assets (1971):	$ 4,350,548	$ 8,052,561
Total Plant Assets (1976):	9,370,507	11,804,140
Grand Total:	$13,721,055	$19,856,701

*adjusted to 1979 dollars @ 8% inflation increase per year

Capital Cost Breakdown for PENELEC/Homer City Plant
(December 1981)

High-Gravity Section: (Medium-Cleaned Coal)	$19,800,000
Low-Gravity Section: (Extensively-Cleaned Coal)	$32,300,000
Modifications:	$38,100,000
Truck Dump:	$ 7,000,000
Total:	$97,200,000

plant completed in 1976. Thus the estimated capital costs of the entire plant ($19.9 million) include the cost of constructing buildings and ancillary facilities (storage and refuse handling facilities), and buying plant equipment for each phase of construction. As can be seen in Table H, the fine coal cleaning plant (buildings and equipment) was more costly than the less sophisticated coarse cleaning plant, even after adjustment for inflation.

Operation and maintenance cost estimates were provided by only four plant owners in three different formats (yearly estimates, costs per ton of raw coal processed, and costs per ton of clean coal processed). The major components of these costs include labor and supervision, electric power, O&M supplies, lubricants, magnetite, reagents and water. Even though there is little basis for comparing O&M costs, PENELEC reported that their plant's $13 million per year O&M costs were high because of the high costs of starting up the preparation plant, and the design-related difficulties that led to more costly O&M.

Coal preparation plants are owned by three general categories of companies: utilities, oil/energy companies and "independent" coal companies. The utilities financed capital expenditures primarily by selling stocks and bonds (e.g., Duquesne Light Company), and by obtaining leases and loans from banks (e.g., PP&L/PMC). The oil/energy companies obtained plant financing from their internal cash flow, which in some cases included the parent company's monies (e.g., Island Creek and R&F Coal Companies). The independent coal companies did not respond to questions concerning the financing of preparation plants.

The only government incentive that companies purchasing preparation plants qualify for is the 10 percent investment tax credit for plant construction.* A number of plant owners believe that other government incentives should exist to aid potential preparation plant builders.

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*Investors in preparation plants are not eligible for pollution control bonds (which have a lower interest rate than regular bonds) as are companies purchasing FGD units.

This Noyes report analyzes the costs of physically and chemically cleaning coal. In addition to the basic capital and annual costs, site-specific and other variables are included to estimate costs more accurately. The economic benefits and penalties of these processes are also described.

Economic Aspects

The information in this chapter is based on:

> *Environmental Assessment of Coal Cleaning Processes: Technology Overview*, EPA Report 600/7-79-073e, September 1979, prepared by P.W. Spaite (Consultant), G.L. Robinson, A.W. Lemmon, Jr., S. Min, and J.H. McCreery of Battelle Columbus Laboratories for the U.S. Environmental Protection Agency.
>
> *Assessment of Coal Cleaning Technology: An Evaluation of Chemical Coal Cleaning Processes*, EPA Report 600/7-78-173a, August 1978, prepared by G.Y. Contos, I.F. Frankel, and L.C. McCandless of Versar, Inc. for the U.S. Environmental Protection Agency.
>
> *Evaluation of Physical/Chemical Coal Cleaning and Flue Gas Desulfurization*, EPA Report 600/7-79-250, November 1979, prepared by T.W. Tarkington, F.M. Kennedy, and J.G. Patterson of the Tennessee Valley Authority for the U.S. Environmental Protection Agency.

Evaluation of the economics of coal cleaning requires consideration of both costs and benefits which are highly specific to individual coal supply/coal user combinations. Further, some of the most important factors are difficult to quantify, e.g., the value of improved boiler operability and efficiency which may result from lower throughput of inert mineral matter and sulfur. Also where coal cleaning is being assessed as an approach to control of sulfur oxide pollution, it must be compared with available alternatives including flue gas desulfurization (FGD), the use of naturally occurring low-sulfur coals, and the use of coal cleaning in combination with FGD. Finally, the economic factors can be assessed from several perspectives, e.g., that of a coal producer evaluating the return on investment that might be realized by selling cleaned coal or that of a coal user evaluating what might be saved using cleaned coal. All of these factors combine to make generalizations difficult. However, one study[1] demonstrates a definite economic advantage for PCC combined with FGD for SO_2 emission control, as compared to the exclusive use of FGD. Other data indicate that coal cleaning can be very attractive economically in many situations, and it can ease the problems in meeting environmental objectives.[2]

PHYSICAL COAL CLEANING COSTS

Only those costs which would be incurred by the operator of a PCC plant are considered here. No attempt is made to estimate the incentives for selling ROM coal, as opposed to physically cleaned coal, or the probable selling price of the latter. The overall cost of coal preparation may be expressed as \$/ton of clean coal, \$/ton of raw coal, $\$/10^6$ Btu, or \$/lb of sulfur removed. Costs are frequently stated in several ways to prevent confusion associated with the use of different terms. The total cost of PCC includes capital charges, operating and maintenance (O&M) costs, and the costs of the Btu lost when combustible matter is rejected in the refuse stream from PCC processes.

Capital Charges

Annualized capital charges depend on capital investment, estimated useful life, L, of the plant (in years) and the interest or discount rate, i (annual percentage). The capital investment required for a PCC plant depends heavily on its size and complexity. The complexity, which depends on the cleanability of the coal and product requirements, is probably the dominant cost-related factor for modern plants. The complexity-cost relationship can be illustrated by considering the following costs (in 1977 dollars): For a simple jig plant processing a 6-inch by 0 coal with the 48 mesh x 0 fraction processed in hydrocyclones to produce a 48 x 120 mesh clean coal fraction, the cost was \$6,600 per ton/hour of feed capacity.[3] For a multistream coal cleaning plant which was built to help control emissions from a power plant, the cost was \$42,000 per ton/hour of capacity.[4]

Capital investment costs are converted to fixed annual charges on capital by estimating L and i. The values of L and i determine the capital recovery factor (CRF) which is multiplied times the capital investment cost to obtain the fixed annual charges on capital. These charges allow for amortization of the capital investment and the time value of the money invested over the useful life of the plant. The approximate PCC capital investment cost range of \$6,600 to \$42,000 per ton/hour of capacity can be converted to a range for charges on capital as follows. With L of 15 years and i of 18% (based on a combination of interest on borrowed funds and return on equity), CRF is 0.19640, which yields fixed annual charges on capital of about \$1,296 to \$8,249 per ton/hour of capacity. With a plant capacity of 1,000 tons/hour and a plant utilization of 42%, or 3,690 hours/year, the charges on capital are about \$0.35 to \$2.24/ton of feed coal or \$0.44 to \$2.79/ton of clean coal if the weight yield is 0.8. (Cost per ton of cleaned coal is obtained by dividing the cost per ton of feed coal by the weight yield.)

Operating and Maintenance Cost

O&M costs for PCC typically include raw materials and supplies, utilities, labor, and overhead. Other costs which may be included are payments to union welfare fund, insurance, payroll taxes, and property taxes. O&M costs (in 1977 dollars) for PCC may range approximately from $0.60[5] to $4.00[6] per ton of feed coal ($0.75 to $5.00 per ton of clean coal if the weight yield is 0.8). The lower figure is for coarse beneficiation, i.e., washing of +⅜-inch material only with the ⅜-inch x 0 fraction being shipped dry as is. The higher figure applies to full beneficiation of a coal with large amounts of -28 mesh material (above 20%) using heavy-medium circuits and full treatment of the fines.

Cost of Btu Loss

The cost of Btu loss is a function of the cost and heating value of the feed coal and the Btu recovery. Formulas for calculating and reporting these costs have been shown by Battelle.[7] The costs for Btu loss as a function of these two variables is shown in Table 23.1.

Table 23.1: Cost of Btu Loss for R = 0.95
$/ton of Clean Coal

F, Cost of Feed Coal ($/ton)	. . . . 100 Y, Weight Yield, Percent.				
	60	70	80	90	95
10	0.83	0.71	0.63	0.56	0.53
15	1.25	1.07	0.94	0.83	0.79
20	1.67	1.43	1.25	1.11	1.05
25	2.08	1.79	1.56	1.39	1.32

Source: EPA-600/7-79-073e

The values in Table 23.1 reflect a cost which assumes that the value of Btu associated with coal in refuse is equal to that of Btu in clean coal. It should be recognized, however, that this is not a true relationship because constituents removed by PCC have a relatively low heating value and, being associated with sulfur compounds and inert mineral matter, they are difficult to utilize effectively. However, calculation of the cost for Btu loss without differentiation with respect to the actual value of Btu in the coal product versus that in the refuse gives a good approximation. It should be noted also that the cost of coal after washing is not simply a function of the cost of input coal and the amount of coal lost. For example, with a $20/ton cost of feed coal and an 80% weight yield, the cost of "raw materials" per ton of clean coal is $25.00. With a Btu recovery of 95% and weight yield of 80%, the cost of Btu loss is $1.25/ton of clean coal as shown in Table 23.1. Part of the increase in raw material of $5.00 is offset by a $3.75/ton increase in the value of the coal product (attributable to its higher heating value) giving a net cost of $1.25 per ton of cleaned coal.

Total PCC Costs

The range in costs previously cited are summarized in Table 23.2.

Table 23.2: Approximate Ranges of PCC Costs

Cost Category	Approximate Cost Range $/ton of Clean Coal
Capital Charges	0.44-2.79
O&M	0.75-5.00
Btu Loss	0.53-2.08
Total	1.72-9.87

Source: EPA 600/7-79-073e

The calculated value of fixed annual charges on capital is very sensitive to the estimated useful plant life and discount rate. Both the capital and O&M costs per ton of feed or clean coal are very sensitive to the estimated plant utilization. Thus, while most plants would fall in the cost range shown, the costs in specific situations are subject to further variation as estimates for these parameters differ from the representative values used here.

Case Studies of PCC Costs

A recent report,[3] based on a study by The Hoffman-Muntner Corporation, presents the results of detailed case studies of the costs of eight PCC plants. These plants represent a spectrum of PCC technology from a relatively simple jig process to relatively complex circuits involving heavy medium cyclones, froth flotation cells, and thermal dryers. These configurations were considered to be typical of those currently in use. The estimated capital investment costs are given in Table 23.3, and the annualized costs, which include operating and maintenance costs, capital charges, and cost of Btu loss, are shown in Table 23.4. The data are all within the ranges shown in Table 23.2 except for the cost of Btu lost in Plant No. 3. For this plant, coal losses combined with high rejection of mineral matter, give a low yield. As a result, the high cost per ton of raw coal, when converted to cost per ton of cleaned coal comes to $4.49/ton, a value of more than twice that for any other plant in the study. This is a good illustration of the degree to which site-specific factors can cause wide variations from any rule of thumb even though it may be generally applicable. The assumptions made in the Hoffman-Muntner study[3] to determine a capital recovery factor varied widely from those used for the Battelle study,[7] but the differences were in large part offsetting and, when adjusted to a common basis, the agreement was good.

Benefits of Physical Coal Cleaning

The benefits associated with the use of physically cleaned coal include the following.

(1) Reduced coal transportation cost

(2) Reduced coal handling and coal pulverization cost

(3) Reduced ash collection and disposal cost

Table 23.3: Description of Coal Preparation Plants[3]

Plant No.	Type of Process	Raw Coal Capacity (TPH)	Complexity	Estimated Capital Investment (mid-1977 $ x 10³)	Btu Recovery	Weight Yield (%)	Moisture Content in Clean Coal	Heating Value of Clean Coal (Moisture-Free) (Btu/lb)	Total Sulfur Content in Raw Coal (%)	Ash Removal*	Sulfur Removal* (lb/T).....
1	Jig	600	Simple	3,946	91.6	59.0	8-9	13,236	0.65	651	3.7
2	Jig	1,000	Intermediate	13,681	96.4	71.4	6.9	12,974	3.42	472	23.3
3	Jig	1,000	Intermediate	12,084	83.0	56.6	4.6	13,056	4.30 (3.01)	606	27.7
4	Jig	1,600	Complex	22,886	93.7	59.6	5.8	14,244	1.0	678	9.0
5	Heavy Medium	1,400 (720)**	Simple	9,962	94.6	74.0	7.5	10,992	4.56	260	29.5
6	Heavy Medium	600	Complex	13,449	89.2	73.3	5.0	14,336	4.38	338	55.2
7	Heavy Medium	600	Complex	8,420	93.1	60.0	4.9	13,348	1.0	660	6.3
8	Heavy Medium	900	Complex	20,916	94.3	86.0	5.0	13,130	2.45	145	18.0

*lb removed/ton of raw coal processed.

**680 TPH of raw coal is not processed by the cleaning plant.

Source: EPA-600/7-79-073e

Table 23.4: Total Annualized Costs for Eight Coal Preparation Plants[3]
(Mid-1977 Dollars)

Plant No.	Operating and Maintenance Cost*		Capital Charges **		Cost of Btu Loss		Total Annualized Cost				
	Raw Coal	Clean Coal	Raw Coal	Clean Coal	Raw Coal	Clean Coal	Raw Coal	Clean Coal	$/10^6 Btu Recovered	Ash Removed	Sulfur Removed
	($/Ton)									($/Ton)	
1	1.59	2.70	0.38	0.65	1.26	2.14	3.23	5.49	0.227	9.92	1,746
2	1.82	2.55	0.80	1.12	0.54	0.75	3.16	4.42	0.183	13.39	271
3	1.51	2.67	0.71	1.25	2.54	4.49	4.76	8.41	0.338	15.71	344
4	1.76	2.96	0.84	1.40	0.95	1.60	3.55	5.96	0.222	10.47	789
5***	2.37	3.20	0.42	0.56	0.81	1.10	3.60	4.86	0.239	27.69	244
6	2.23	3.04	1.31	1.79	1.62	2.21	5.16	7.04	0.258	30.53	187
7	1.27	2.12	0.82	1.36	1.06	1.76	3.15	5.24	0.206	9.55	1,000
8	2.10	2.44	0.81†	0.94†	0.88	1.02	3.79	4.40	0.176	52.28	421

*Includes labor, supervision, overhead, supplies, fuel, electricity, and subcontract services.

**Based on a 10-year amortization period, 9% discount rate, and 30% utilization factor, except as noted.

***Costs shown for Plant No. 5 are based on 1,400 TPH.

†50% utilization factor.

Source: EPA-600/7-79-073e

(4) Increased boiler efficiency, capacity, and availability
(5) Reduced O&M costs
(6) Savings in cost for control of sulfur oxide pollution

Items (1) through (5) above are savings in operating cost which result from general upgrading of coal quality. Item (6) is a savings which is realized where PCC is used to minimize sulfur pollution control costs. To date there has been no comprehensive assessment of the value of all these benefits. A number of relevant studies have been conducted, however, and some understanding of the potential importance of some of the factors above is being developed.[6]

CHEMICAL COAL CLEANING COSTS

This section presents a comparison of technical and economic results obtained from the assessment of major chemical coal cleaning processes. The analysis and conclusions presented herein are based on process claims made by individual developers, research reports and published information.

Most of the processes are not at an adequate developmental stage to permit the preparation of a precise engineering process flow sheet for capital and operating cost evaluation. Thus, the process economics presented for most processes are best engineering estimates, based upon the information available.

Sulfur Removal and Heating Value Recovery Potential

A comparison of process performance and costs can best be accomplished by looking at each process on a common coal feed. This basis allows the comparison of the following parameters process by process:

- Weight yield of clean coal product based upon a feed coal rate (moisture-free basis) of 7,110 metric tons (7,840 tons) per day [7,200 metric tons (8,000 tons) per day of 2% moisture coal];
- Weight percent sulfur in the clean coal product based upon the sulfur removal efficiency of the process;
- Heating value yield of the process based upon a feed coal value of 6,800 kg cal/kg (12,300 Btu/lb) and net heating value yield in percent; and
- Costs–
  - Total capital costs for the process;
  - Total annual processing costs;
  - Annual costs per metric ton of clean coal, including coal costs and excluding coal costs; and
  - Annual costs per heating value unit, including coal costs and excluding coal costs.

This comparison data is shown in Table 23.5, arranged according to categories of processes.

The common coal feed selected is a bituminous coal from the Pittsburgh seam, which cannot readily be cleaned by conventional washing techniques to meet the current new source performance standards for sulfur dioxide emission. This coal does have an organic sulfur content low enough (0.7 wt %) so that complete removal of pyritic sulfur would result in a product which will meet current NSPS for sulfur dioxide emission.

The percent removal of pyritic and organic sulfur assigned to each process is based on data supplied by individual developers. The table indicates a range of SO_2 emission levels for the clean coal products of 1.5 to 3.8 kg/10^6 kg cal (0.8 to 2.1 lb/10^6 Btu). The calculated sulfur levels for processes which remove both types of sulfur are lower than the 2.2 kg/10^6 kg cal (1.2 lb/10^6 Btu) NSPS for sulfur dioxide emission. Of the four processes which remove pyritic sulfur only, two (TRW and Ledgemont) will produce slightly higher sulfur levels than that required to meet the current NSPS; however, within the levels of accuracies involved they also might be considered to be in compliance. The remaining two processes (Magnex and Syracuse) would produce coal which would be in compliance only with a standard of 4.3 kg/10^6 kg cal (2.4 lb/10^6 Btu) for sulfur dioxide emission.

Processes which remove pyritic sulfur alone are primarily applicable to coals rich in pyritic sulfur, so that efficient removal of pyritic sulfur could bring these coals into compliance. Processes which remove both types of sulfur are primarily applicable to coals which cannot be adequately treated by pyritic removal processes. All chemical coal cleaning processes are more selective and efficient than conventional coal cleaning techniques and it is very likely that each process will eventually find an area of application.

As shown in Table 23.5, the heating value yields estimated for these processes are generally greater than 90% with a range from a low 57% for the IGT process to a high of 96% for the GE process. All heating value yields listed in Table 23.5 reflect both the coal loss due to processing and the coal used to provide in-process heating needs. However, with the exception of the IGT process, the actual coal loss due to processing is claimed to be small. For most processes, the major heating value loss is due to the use of clean coal for in-process heating.

It is believed that the high yield estimated for the GE process may not adequately reflect the heat requirements that may be needed to regenerate the caustic reagent employed in the process. This process is in its early stage of development and as such, the energy requirements for the process cannot be properly assessed at this time. It is possible, that in the final analysis, the heating value recovery from this process will be more in line with other chemical coal cleaning processes.

Table 23.5: Process Performance and Cost Comparison for Major Chemical Coal Cleaning Processes

		Processes Which Remove Pyritic Sulfur Only				Processes Which Remove Pyritic and Organic Sulfur						
	Feed*	TRW	LOL	Magnex	Syracuse Physical Cleaning	ERDA	GE	Battelle	JPL	IGT	KVB	ARCO
Net coal yield,**												
Metric tons/day	7,110	6,400	6,400	5,645	5,645	6,400	6,826	6,755	6,470	4,270	6,070	6,400
Tons/day	7,840	7,056	7,056	6,225	6,225	7,056	7,526	7,448	7,135	4,704	6,690	7,056
Sulfur in the product, wt %	1.93	0.83	0.83	0.97	1.50	0.65	0.50	0.65	0.6	0.55	0.61	0.69
Heating value,												
kg cal/kg	6,800	7,100	7,100	6,900	8,100	7,100	6,800	6,300	6,800	6,500	7,300	6,900
Btu/lb	12,300	12,835	12,835	12,400	14,600	12,835	12,300	11,350	12,300	11,685	13,120	12,400
kg SO_2/MM kg cal	5.6	2.3	2.3	2.8	3.8	1.6	1.5	2.2	1.8	1.6	1.6	2.0
lb SO_2/MM Btu	3.1	1.3	1.3	1.6	2.1	0.9	0.8	1.2	1.0	0.9	0.9	1.1
Net Btu yield, %	–	94	94	80	95	94	96	88	91	57	91	91
Costs												
Capital, $MM	–	109	114.0	37.8	49.0	166.8	102.0	168.6	103.2	134.6	65.9	–
Annual processing, $MM	–	37.2	45.3	19.2	12.2	56.6	39.8	74.2	44.4	38.3	41.0	58.7
Clean coal, excluding coal costs												
$/annual metric ton	–	17.2	21.0	10.1	6.4	26.3	17.3	32.6	20.4	26.6	20.1	27.5
$/annual ton	–	15.6	19.1	9.2	5.8	23.8	15.7	29.6	18.5	24.2	18.2	25
Clean coal, including coal cost***												
$/annual metric ton	–	47.9	51.6	44.8	40.8	56.9	46.0	61.6	50.7	72.6	52.4	–
$/annual ton	–	43.4	46.9	40.7	37.0	51.6	41.8	55.9	46.0	65.8	47.5	–
Excluding coal cost												
$/MM kg cal	–	2.42	2.94	1.47	0.78	3.69	2.54	5.15	2.97	4.09	2.75	–
$/MM Btu	–	0.61	0.74	0.37	0.20	0.92	0.64	1.30	0.75	1.03	0.69	–
Including coal cost***												
$/MM kg cal	–	6.71	7.23	6.52	5.03	7.98	6.72	9.74	7.40	11.2	7.22	–
$/MM Btu	–	1.69	1.82	1.64	1.27	2.00	1.69	2.45	1.86	2.81	1.81	–

*The coal selected is a Pittsburgh seam coal from Pennsylvania which contains 1.22 wt % pyritic, 0.01% sulfate and 0.70% organic sulfur. It is assumed that this coal has a heating value of 6,800 kg cal/kg (12,300 Btu/lb).

**All values reported are on a moisture-free basis.

***Assumes coal feed at $27.6/metric ton ($25/ton).

Source: EPA 600/7-78-173a

Cost Comparison for Major CCC Processes

Estimates of capital and annual operating costs for each major chemical coal cleaning process are also given in Table 23.5. These estimates are based on an assumed plant throughput capacity of 7,200 metric tons (8,000 tons) per day, equivalent to a 750 MW electric power plant. The total annual operating costs for each process, including and excluding cost of the raw coal, have been expressed also in terms of dollars per metric ton and dollars per million kg cal heat content in the coal.

The capital cost estimate prepared by each process developer was used as the basis of the cost estimates in this report. In some cases, these costs were modified to allow the evaluation of the various processes on a comparable basis. The estimated capital costs assume a grass roots operation including costs for coal crushing, grinding, product compacting and feed and product handling. The capital costs also include land acquisition and site development, off-site facilities, and engineering and design costs. A contingency allowance of 20% has been included in all estimates, with the exception of TRW's. A lower contingency allowance (10%) was used for the TRW process since it is at a more advanced stage of development and adequate process data is available to develop the economics of this process with a greater degree of confidence.

Annual operating costs are based on a 24-hour workday, 90.4% service factor (330 days per year) basis. The capital cost is amortized over a period of 20 years at 10% interest per year. Where adequate information was available, the utilities and chemical consumptions are based upon actual process demand. The operating labor costs reflect wage rates for the Pittsburgh, Pennsylvania, area. The estimates for the maintenance and supplies, general and administrative, taxes and insurance are taken as 5, 1.5, 2 and 1% on total installed plant capital cost (TPC), respectively.

Capital Cost Comparisons: In general, pyritic sulfur removal processes require the least amount of capital investment. However, these processes have limited sulfur removal efficiencies.

Among processes that remove both organic and pyritic sulfur, the KVB process appears to have the lowest capital investment, since it is a partially dry process requiring lower investment for the dry reaction section. The high capital cost of the Battelle process is due to the processing steps associated with reagent regeneration.

The high capital cost of the ERDA process is due to costly equipment associated with the handling of dilute sulfuric acid at elevated temperatures and pressures. At the process operating conditions the dilute acid is highly corrosive and it poses problems in terms of selection of construction material for equipment and devices which are exposed to the corrosive atmosphere.

Very little is known about the ARCO process details and process chemistry. Therefore, a capital cost estimate was not developed for that process.

Operating Cost Comparisons: Table 23.6 presents a summary of operating cost elements for each process. The ranges of annual operating costs, including raw coal cost, in terms of \$/metric ton and \$/10^6 kg cal are \$43.40 to \$72.40 and \$5.38 to \$11.20, respectively. Pyritic sulfur removal processes using chemical pretreatment are the least expensive of all processes listed in Table 23.6. Operating cost for the Magnex process depends primarily on the cost of iron pentacarbonyl manufacturing. In the estimate presented in Table 23.5, an operating cost of \$0.22/kg for the iron carbonyl manufacturing was used, as projected by the developer. The current vendor quotes for iron carbonyl range up to \$3.30/kg. At a consumption rate of 10 kg/metric ton of coal, each \$0.20 cost increase per kilogram of iron carbonyl manufactured would increase the annual operating cost of this process by about 27%.

Between the two processes which remove pyritic sulfur by leaching, the TRW process appears to be slightly less costly. In the Ledgemont process the fixed charges associated with the higher capital investment have an adverse impact on the annual operating costs. Additionally, the TRW process has a much higher probability of technical success since it is currently active at a PDU stage. The Ledgemont process, tested only at a minipilot plant level, is currently inactive.

The most expensive processes, in terms of energy output, are the IGT process followed closely by the Battelle process. Available laboratory data indicate a very low Btu recovery for the IGT process. The Battelle process is adversely impacted by the fixed charges associated with the high capital investment and by the costs associated with chemicals consumption and reagent regeneration operations.

The least expensive process capable of removing pyritic and organic sulfur is the GE process followed closely by the JPL and KVB processes. The GE estimate is based, however, on early laboratory data and it is quite possible that the projected costs will prove somewhat inaccurate in the long run. The basic process utilizes a caustic reagent in coal pretreatment and the costs associated with caustic consumption and caustic regeneration are questionable at this time. The JPL process estimates are also preliminary. The annual costs reported for the KVB process are also preliminary since the process is at its early stages of development and accurate conceptualization of the process for purposes of economic evaluation is not possible. The main advantage of the KVB process is the simplicity of the first stage dry oxidation process. If the dry oxidation process can be successfully demonstrated using coarse coals, this process would be an inexpensive technology for beneficiation of coals where partial removal of sulfur would substantially upgrade the coal.

Among the processes capable of removing pyritic and organic sulfur the ERDA process has one of the highest probabilities of technical success. The process is currently active and most technologies employed in this system have been already tested in other systems such as Ledgemont and TRW. The process is attractive because it is claimed to remove both types of sulfur and uses air

Table 23.6: Operating Cost Comparisons for Major Chemical Coal Cleaning Processes

	Process									
	Meyers		Ledgemont		Magnex		Syracuse		ERDA	
Cost Element	Cost ($1,000)	$/Ton Product	Cost ($1,000)	$/Ton Product	Cost ($1,000)	$/Ton Product	Cost ($1,000)	$/Ton Product	Cost ($1,000)	$/Ton Product
Labor & G & A	3,962	1.67	1,600	0.67	786	0.38	620	0.31	3,255	1.36
Amortization (20 yr)	12,820	5.39	13,400	5.64	4,444	2.12	5,750	2.72	19,600	8.25
Taxes and Insurance	3,270	1.38	3,400	1.44	1,135	0.54	1,470	0.69	5,004	2.11
Maintenance and Supplies	5,460	2.29	7,300	3.07	1,891	0.90	3,170	1.50	8,340	3.51
Utilities	5,764	2.43	10,600	4.46	1,400	0.67	140	0.06	13,224	5.57
Chemicals	4,692	1.97	8,200	3.45	9,144	4.36	4,220	1.99	6,932	2.92
Waste Disposal	1,275	0.54	800	0.34	438	0.21	–	–	240	0.10
Annual Processing Cost	37,243	15.67	45,300	19.07	19,238	9.18	12,190	5.77	56,595	23.82
Raw Coal	66,000	27.78	66,000	27.78	66,000	31.49	66,000	31.25	66,000	27.78
Total Annual Cost	103,243	43.45	111,300	46.85	85,238	40.67	78,190	37.02	122,595	51.60
	GE		Battelle		JPL		IGT		KVB	
Labor & G & A	1,830	0.72	4,630	1.85	3,700	1.54	5,095	3.22	1,429	0.63
Amortization (20 yr)	11,980	4.73	19,814	7.90	7,274	3.03	15,800	9.97	7,748	3.44
Taxes and Insurance	3,060	1.21	5,050	2.00	1.326	0.55	4,050	2.58	1,978	0.88
Maintenance and Supplies	5,310	2.10	10,000	3.99	2,210	0.92	6,732	4.24	3,297	1.46
Utilities	7,270	2.87	23,100	9.21	1,300	0.54	3,300	2.08	17,271	7.67
Chemicals	10,370	4.09	11,609	4.63	28,600	11.91	3,300	2.08	9,204	4.09
Waste Disposal	–	–	–	–	–	–	–	–	131	0.06
Annual Processing Cost	39,820	15.72	74,203	29.58	44,410	18.49	38,277	24.17	41,059	18.23
Raw Coal	66,000	26.05	66,000	26.32	66,000	27.48	66,000	41.66	66,000	29.31
Total Annual Cost	105,820	41.77	140,203	55.90	110,410	45.97	104,277	65.83	107,059	47.54

Source: EPA 600/7-78-173a

Table 23.7: Cost Effectiveness and Other Comparisons of Chemical Coal Cleaning Processes*

Process	Type of Sulfur Removed	Sulfur in Product* (%)	Sulfur Removed (%)***	Process Cost ($/metric ton including cost of coal)	Cost Effectiveness of S Removal ($/% S removed)	Cost Effectiveness Ranking§§	Meets EPA NSPS*	Probability of Success (%)	Time Frame for Commercial Availability (Years)†
Magnex	P**	0.97	0.96	44.8	46.6	2	No	85	2-3
Syracuse and Physical Cleaning	P	1.50§	0.43	40.8	94.9	4	No	70	2-3
TRW	P	0.83	1.10	47.9	43.5	1	No	90	<3
LOL†††	P	0.83	1.10	51.6	46.9	3	No	50	4-5
ERDA	P&O***	0.65	1.28	56.9	44.5	4	Yes	70	5
GE	P&O	0.50	1.43	46.0	32.2	1	Yes	60	5
Battelle	P&O	0.65	1.28	61.6	48.1	5	Yes	35	4-5
JPL	P&O	0.60	1.33	50.7	38.1	2	Yes	55	5
IGT	P&O	0.55	1.38	72.6	52.6	6	Yes	20	5
KVB	P&O	0.68	1.25	52.4	41.9	3	Yes	10†††	5
ARCO	P&O	0.69	1.24	–††	–††	–	Yes	–††	–††

*Based on Pittsburgh seam coal from Pennsylvania which contains 1.22 wt % pyritic, 0.01 % sulfate and 0.70% organic sulfur.

**P = pyritic sulfur.

***P&O = pyritic and organic sulfur.

†Time frame assumes continuing effort or renewed effort starting immediately.

††Information available is insufficient to make educated guesses.

†††Processes not currently active, partially accounting for low probability of success.

§80% yield of product assumed in cleaning plant.

§§Based on type of sulfur removed.

Source: EPA 600/7-78-173a

as a major reagent. Furthermore, the sulfur by-product from this process is claimed to be dilute sulfuric acid, rather than iron sulfate, which greatly simplifies the coal washing operations. The process is somewhat expensive due to high operating temperature and pressure requirements and the corrosive nature of dilute acid present in this system. The dilute sulfuric acid at the operating conditions of the ERDA process will require the use of expensive construction material and consequently a higher capital investment cost.

Table 23.7 presents a cost effectiveness summary derived from information presented in Table 23.5. Costs are presented in terms of dollars per percent of sulfur removed from coal regardless of the quality of the treated product. However, column 7 of the table shows whether the product would comply with the current EPA's NSPS for SO_2 emissions. The processes are then rated based upon the cost effectiveness of sulfur removal. The subjective probability of success assigned to each process shown in column 8 of this table is based on integration of several factors such as:

- available experimental data;
- status of the process;
- known product quality deficiencies;
- known process problems; and
- the degree and quality of effort assigned to the individual program.

In conclusion, all chemical coal cleaning processes discussed in this section offer a possibility of converting coal into clean fuel. Each process has an area of application. However, processes that remove both pyritic and organic sulfur will have a greater impact in coal utilization. If chemical coal cleaning is to be used as an approach for greater utilization of coal as an environmentally acceptable fuel, the pyritic and organic sulfur removal processes should be given the most emphasis and support.

OTHER ECONOMIC CONSIDERATIONS

Site-Specific Cost Factors

It should be emphasized that many site-specific variables affect the economics involved in determining the lowest cost method to meet emission standards. Some of these are discussed below.

Emission Standards: Costs for coal cleaning, alone or in combination with FGD, will be most attractive for plants under higher State Implementation Plan (SIP) standards, less attractive for 1.2 lb SO_2/MBtu emission limits, and least attractive for 85% SO_2 removal, as compared with FGD alone. Raising the proposed 0.2 lb SO_2/MBtu floor in the 85% SO_2 removal NSPS would have increased the number of situations in which coal cleaning would be economically attractive for that standard.

Coal Properties: Higher ratios of pyritic sulfur to organic sulfur will result in improved costs for most situations utilizing coal cleaning. Larger pyrite crystal size or more favorable particle distribution will make the pyrite easier to remove by PCC and will also result in improved costs. Data exist which show that coal cleaning will not only reduce the sulfur level, but also the sulfur variability in coal. Coal cleaning plus FGD will generally have more attractive costs for a raw coal with high sulfur variability. The size and cost of a coal cleaning plant to deliver a specified heat load are also obviously dependent on the heat content of the raw coal.

Plant Location: Plant location influences costs in a variety of ways. These include construction and operating labor rates; need for weather protection; requirements for barge, truck, or rail unloading facilities; size of raw and cleaned coal storage facilities; and rates for utilities. If the coal cleaning plant is located adjacent to the power plant, the cleaned coal could be considered as part of the power plant normal 90- to 120-day stockpile. This would reduce both the installed capital and working capital.

Process Commercial Development: The PCC processes are in a commercial stage of development whereas the CCC processes are all relatively undeveloped. Future development work on the CCC processes could substantially increase or decrease the projected costs. For example, the significant reduction in sodium hydroxide usage recently reported by KVB for their process would make it more economically attractive.

While the use of absorbents such as limestone to remove SO_2 during combustion has not been fully successful in the past, in recent work reported by Battelle burning pellets made of finely ground limestone and coal showed a sulfur capture of about 75% in the boiler. After cleaning, the cleaned coal has a fine particle size and in some cases it must be pelletized for shipping or storage. These grinding and pelletizing or briquetting costs are included in the costs of the CCC processes. Grinding limestone and pelletizing it with the cleaned coal, using cement as a binder, would offer many process and cost advantages for both SO_2 control methods. For many coals this technique, if perfected, used with coals cleaned by the KVB process could even meet 85% SO_2 removal standards.

Other Economic Benefits and Penalties of Using Cleaned Coal

In evaluating the capital investment and annual revenue requirements associated with coal cleaning, it is useful to also assess the other economic benefits and penalties that result from use of cleaned coal. In addition to the primary benefit that the cleaned coal is lower in sulfur, it is generally also lower in ash and higher in heating value, although often higher in surface moisture. Combustion of coal with these characteristics has numerous benefits as well as certain disadvantages to the user. The net effect is a credit which may be of sufficient magnitude to offset some of the increased cost of cleaned coal.

Transportation Costs: Coal beneficiation at the mine decreases the cost of coal transportation by increasing the heating value of the coal, consequently reducing the quantity of coal necessary to supply a given heat requirement.

Pension and Benefit Trust Fund: Provisions of the 1978 UMW contract require payment by the mine operator of $1.385 to the UMW Pension and Benefit Trust Fund for each ton of coal shipped to a consumer. Less cleaned coal is needed to supply a specified heat load requirement and, if the coal cleaning plant is at the mine, the reduced tonnage will decrease this payment.

Pulverization Costs: PCC, by reducing mineral matter, decreases coal hardness and facilitates crushing. The increased heating value of cleaned coal also reduces the quantity of coal to be crushed. The size of the cleaned coal product is considerably smaller than that of raw coal so that significant pulverization costs, which are already covered in the coal cleaning costs, are saved. Additional surface moisture of cleaned coal may partially offset these advantages.

Boiler Capacity: The higher heating value of cleaned coal decreases the possibility that the utility boiler capacity will be derated because of deteriorating coal quality. Also, by reducing the slagging tendency of the coal, coal cleaning can permit the design of furnaces with higher heat transfer rates and correspondingly smaller furnace volume.

Boiler Performance: Cleaned coal can improve boiler performance by reducing slagging, fouling, and corrosion problems. This can significantly reduce the cost of boiler operation and maintenance and increase the availability of the generating facility.

Ash Handling: Ash handling and disposal costs are decreased since coal cleaning generally reduces the total amount of ash handled. Less sensible heat is lost in the bottom ash because of the lower ash levels.

FGD Operation: FGD systems generally have markedly better operation with low-sulfur coals. When coal cleaning is followed by FGD scrubbing, the lower sulfur level of the cleaned coal should give less FGD system downtime and a better overall utility availability. Better overall utility availability is also obtained with cleaned coal because of the more consistent SO_2 inlet concentration to the FGD system.

FGD Capital and Operating Costs: Investment and operating costs of FGD systems are proportional to inlet sulfur concentration. Boilers burning high-sulfur coal have higher capital costs because of the necessity for a large absorbent preparation facility, scrubber system, and area for waste disposal. Corresponding savings in operating costs should also be realized with the low-sulfur cleaned coal, particularly if partial scrubbing can replace full scrubbing.

Electrostatic Precipitator Size and Cost: The resistivity of fly ash is a major factor in determining the collection area of the ESP. Resistivity is determined by many factors, including ash composition and concentration and SO_3 level in the gas. With conventional ESP units the removal of fly ash will generally be more difficult with the low sulfur levels of cleaned coal. ESP costs may be increased although ash levels may be partly compensating. Other systems such as hot side ESP, bag filters, or pulsed ESP may be less expensive in certain cases when burning cleaned coal.

Surface Moisture: Higher moisture levels in cleaned coal resulting from the smaller particle sizes increase transportation costs and result in a minor heat loss when the water is heated and vaporized during the combustion process.

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(7) Hall, E.H., Lemmon, A.W., Jr., Robinson, G.L., Goodman, F.K., McCreery, J.H., Thomas, R.E. and Smith, P., "The Use of Coal Cleaning for Compliance with SO_2 Emission Regulations," draft report from Battelle's Columbus Laboratories to U.S. EPA (August 1979).

This National Academy of Sciences study focuses on FGD systems, dividing them into throwaway and regenerable processes, and wet and dry systems. Under each heading, a description of the process is given. Special attention is focused on waste production, and available methods for its disposal.

CHAPTER 4

FLUE-GAS DESULFURIZATION TECHNOLOGY AND PROCESSES

DESULFURIZATION PROCESSES

More than 100 processes and/or systems have received some level of consideration for use in removing the sulfur from flue gases. However, only three or four of the process concepts are now commercially important in the United States. These processes all use aqueous scrubbing with disposal of a waste product or with regeneration of the scrubbing agent to produce marketable sulfuric acid or elemental sulfur. Dry processes have not yet received any significant degree of commercial acceptance, although several types are now being introduced commercially.

Throwaway Processes

The most generally accepted approach to flue-gas desulfurization is aqueous scrubbing with the production of $CaSO_3$-$CaSO_4$ solids. This process generally is known as throwaway scrubbing. Limestone ($CaCO_3$) or lime (CaO) reacts with the flue gas in an aqueous system to produce $CaSO_3$. Because most waste gases contain oxygen, the solids will also contain variable quantities of $CaSO_4$.

Intentional air oxidation of the solids produced has been practiced in Japan in order to produce marketable gypsum, and in the United States it has been tested with a view to improving the disposal characteristics of the solids. Generally, in the United States the $CaSO_3$-$CaSO_4$ or $CaSO_4$ solids are disposed of in ponds or landfills. The disposal of this solid waste in an acceptable manner is one of the most significant problems of throwaway scrubbing.

Slurry Scrubbing. Slurry scrubbing is the dominant commercial FGD process. The general concept is shown in Figure 4-1. It includes a scrubber, a recycle tank, and a device for liquid-solid separation.

Figure 4-1. Throwaway slurry scrubbing.

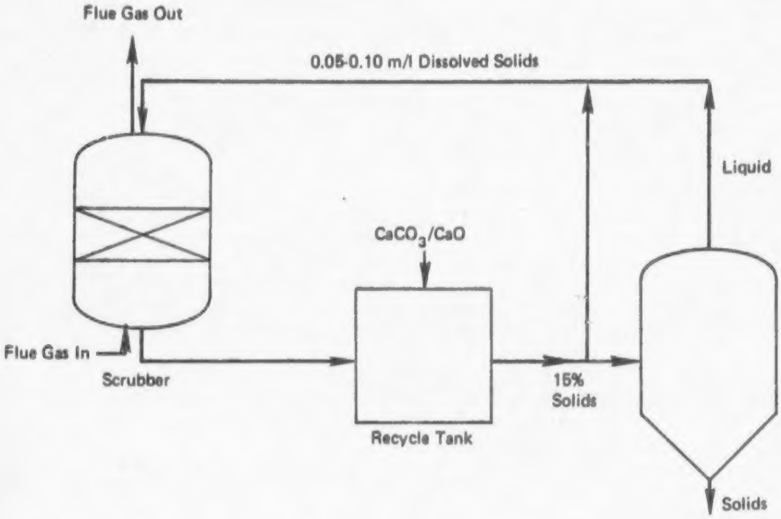
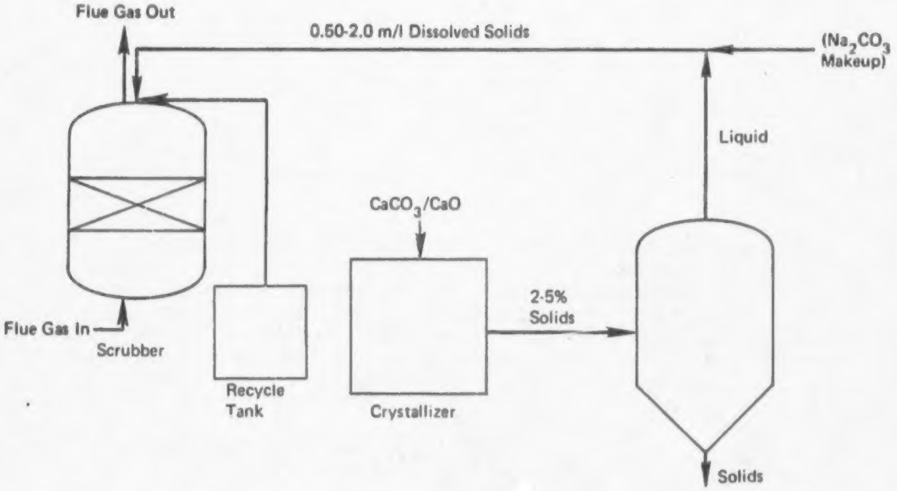

Flue Gas Out
0.05-0.10 m/l Dissolved Solids
Liquid
$CaCO_3/CaO$
Flue Gas In
Scrubber
15% Solids
Recycle Tank
Solids

Figure 4-2. Dual alkali throwaway scrubbing.


Flue Gas Out
0.50-2.0 m/l Dissolved Solids
(Na_2CO_3 Makeup)
Liquid
$CaCO_3/CaO$
2-5% Solids
Flue Gas In
Scrubber
Recycle Tank
Crystallizer
Solids

Typically, the scrubber operates at the adiabatic saturation temperature (50-60°C for most flue gas) and atmospheric pressure. Clean gas from the scrubber is usually reheated to 75-90°C by hot-air injection or indirect exchange with steam to allow for adequate plume dispersion and to protect downstream equipment from corrosion by acidic condensate. A number of scrubber types have been used, including sprays, venturis, turbulent beds of solid or hollow spheres, modified sieve trays, and open eggcrate packing.

The recycle tank is usually a simple stirred tank or several tanks in series. Slurry is recycled to the scrubber, with a small portion, containing 10 to 15 percent (by weight) suspended solids, withdrawn for liquid-solid separation. Clarifiers, filters, centrifuges, and settling ponds have all been used for liquid-solid separation.

The design of this type of system is constrained primarily by the need to avoid $CaSO_3$ and $CaSO_4$ scaling in the scrubber. To control $CaSO_4$ scaling, a combination of crystallizer residence time and $CaSO_4$ solids concentration is used to desupersaturate the scrubber feed. The liquid's circulation rate must also be adequate to prevent excessive $CaSO_4$ supersaturation at the scrubber exit. $CaSO_3$ scaling is prevented by limiting the amount of unreacted $CaCO_3$ or CaO solid that is returned to the scrubber.

If the amount of sulfate in the solids is less than 15 to 20 percent of the total sulfate plus sulfite, $CaSO_4$ will crystallize in solid solution with $CaSO_3$, rather than crystallizing as gypsum ($CaSO_4 \cdot 2H_2O$), its usual form.[2 3] As a result, the constraints on crystallizer design and liquid circulation rate can be relaxed without the occurrence of gypsum scaling.

Soluble additives, such as MgO, Na_2CO_3, and organic acids, have been used to enhance the mass transfer performance of slurry scrubbers.[2 4-6] Air oxidation has been used to upgrade the waste solids (because gypsum is easier to dewater than $CaSO_3$).[7 8]

Air oxidation, limestone dissolution, and gypsum crystallization all take place within the scrubber vessel in the Chiyoda Thoroughbred 121 system, which uses a sparger (gas-phase-dispersed) contactor as the scrubber instead of the liquid-phase-dispersed contactors described above. The Chiyoda system has been operated at a 23-MW prototype test unit on a medium-sulfur coal-fired utility boiler in the United States.

Clear-Solution Scrubbing. Clear-solution scrubbing avoids many of the problems associated with slurry scrubbing.[1] One such process is shown in Figure 4-2. This dual alkali process absorbs SO_2 in a clear solution of sodium sulfite (Na_2SO_3) to produce sodium bisulfite ($NaHSO_3$). The scrubbing solution is regenerated by lime or limestone with the resulting crystallization of $CaSO_3$ solids.[9-11]

If less than 15 to 20 percent of the SO_2 is oxidized in the scrubber system, $CaSO_4$ is precipitated in solid solution with the $CaSO_3$.[12] Otherwise, the sulfate must be removed as Na_2SO_4 solids or solution, although some Na_2SO_4 can be regenerated with CaO to yield gypsum.

The $CaSO_3$-$CaSO_4$ solids are separated from the scrubbing solution by clarification followed by filtration or centrifugation. The filter cake is washed with makeup water to retain soluble sodium in the scrubbing system and thereby minimize sodium makeup.

The clear-solution dual alkali technology is now being commercialized because it is more efficient and may be more reliable than slurry scrubbing. However, the liquid-solid separation system is more complex, and only lime has been used successfully as the alkali source.

Another approach to clear-solution scrubbing is represented by the Dowa process.[13] Sulfur dioxide is absorbed in an aqueous solution of basic aluminum sulfate buffered at pH 3 to 4. After every pass through the scrubbing system, the solution is oxidized by air to convert the dissolved SO_2 to sulfate. A slipstream from the scrubber loop is neutralized by lime or limestone, with a resultant crystallization of gypsum, which is separated by a centrifuge. This process has been commercialized in Japan, and there are plans to begin testing it in the United States in 1980 at the TVA Shawnee Power Plant.

Dry Scrubbing. Dry scrubbing processes that utilize reactive solids (generally lime, sodium carbonate, or sodium bicarbonate) have been studied for a number of years. Early efforts, where lime or limestone was injected directly into the gas stream, gave low efficiencies and poor utilization of reactant. Sodium compounds gave better efficiency and utilization but were costly and resulted in formation of soluble solid waste that could present environmental problems. Recent technical advances, combined with new air emission standards that require cleaning stack gases with relatively low concentrations of SO_2, have generated renewed interest in dry scrubbing systems.

Several versions of dry scrubbing have emerged. These involve the following:

1. Dry reactant, injected directly into the gas stream for reaction, followed by collection and further reaction on

a fabric filter or collection in an electrostatic precipitator. This process is being installed on full-scale equipment.

2. Pelletizing coal with powdered limestone prior to combustion so that reaction can take place during combustion, in the flue gas, and during the dust collection process. This process has shown promise on pilot-scale equipment.

3. Spray drying, utilizing a spray dryer to contact Na_2CO_3 solution or CaO slurry with the stack gas. The solution or slurry is completely evaporated (no liquid is present at the outlet) and the dry solids are collected by a fabric filter or electrostatic precipitator. As a result, the use of corrosion-resistant materials and liquid recirculation is generally not required and the reheat requirement is minimized. This process is being installed on full-scale equipment.

It is expected that this technology will be limited to stack gases with low levels of SO_2 (less than 1000 ppm). At higher SO_2 concentrations, lime scrubbing systems have been unable to achieve 90 percent SO_2 removal; Na_2CO_3 becomes prohibitively expensive and environmentally unacceptable because of the large amount of alkali consumed and soluble salt waste produced.

Regenerable Scrubbing Processes

Regenerable processes remove SO_2 from flue gas and convert it to marketable products, such as elemental sulfur, sulfuric acid, and liquid sulfur dioxide. Such systems are environmentally attractive because they minimize greatly the production of waste products. Economically, these systems are potentially attractive because they avoid the costs of alkali makeup and solid waste disposal. In practice, the energy required for processing and the complexity of the processing usually make these processes more expensive than throwaway scrubbing. As a result, there are now only a few applications of regenerable processes in the United States, although there continues to be an active interest in developing such processes. Regenerable processes are more commonly used in Japan because waste disposal is more of a problem there than in the United States.

As long as most facilities continue to use throwaway FGD systems, there will be adequate markets for the sulfur and sulfuric acid produced in regenerable processes in the United States.[14] [15] Sulfur is the preferred product because it is easily stored and inexpensively transported. If there are no markets, sulfur is also an acceptable waste product, since it can be stored easily. However, production of

Table 4-1. Regenerable scrubbing processes.

Generic Name	Process Operation	Product
Thermal regeneration		
Wellman-Lord[16,17]	Na_2SO_3 solution; evaporative crystallization	95% SO_2
Steam stripping[18,19]	Acid/base buffer or nonvolatile aldehyde	95% SO_2
MgO[20-23]	MgO/$MgSO_3$ slurry; centrifuge, dry, and calcine	15% SO_2
ZnO[6]	ZnO/Na_2SO_3 solution; centrifuge, dry, and calcine	95% SO_2
Acid decomposition of bisulfate		
Stone and Webster/Ionics[25]	NaOH solution; electrolytic prod. of NaOH/H_2SO_4	95% SO_2
Ammonium bisulfate[32,33]	$(NH_4)_2SO_3$ solution; thermal decomp. of $(NH_4)_2SO_4$ to NH_3 and NH_4HSO_4	95% SO_2
H_2S regeneration to S[34,53]		
Citrate[35,36]	Sodium citrate buffer	S
Aquaclaus[37,38]	Sodium phosphate buffer	S
NH_3-IFP[38,39]	$(NH_4)_2SO_3$ solution; evaporative decomp.; aqueous reaction of H_2S with SO_2	S
CO regeneration		
Consol[40]	K_2SO_3 solution reaction with CO	50% H_2S
Sulfoxel	Na_2SO_3 solution reaction with CO	S
High T reduction		
Kel-S[38,41]	$CaCO_3$ slurry; reaction with coal to CaS; carbonation	H_2S
Aqueous carbonate[42,43]	Na_2CO_3 solution; reduction of molten Na_2SO_4 and Na_2SO_3 with coal to make Na_2S; carbonation	H_2S
RESOX[38,46]	Reduction of SO_2, etc.	S

sulfur from SO_2 requires the use of a reducing agent, usually H_2 made from methane. As a result, it is usually more economical to produce sulfuric acid, rather than sulfur, even though sulfuric acid can be three times as costly to transport. There are only limited existing markets for liquid SO_2, but if a large source of SO_2 can be developed, SO_2 has the potential to be substituted for elemental sulfur in the production of sulfuric acid.

A large number of regenerable scrubbing processes have been proposed and partially developed, as shown in Table 4-1. Only the Wellman-Lord and MgO processes are now used commercially. The federal government is currently cofunding large-scale demonstrations (100 MW) of the Wellman-Lord, citrate, and aqueous carbonate processes, and there are active pilot plant and laboratory development programs of the steam stripping, RESOX, and Kel-S processes.

Wellman-Lord Process. The Wellman-Lord process (Figure 4-3) is the only regenerable system currently being used in the United States.[16] [17] In this process, SO_2 is absorbed in a concentrated aqueous solution of sodium sulfite (Na_2SO_3) to produce sodium bisulfite ($NaHSO_3$). The scrubbing solution is regenerated by crystallization of Na_2SO_3 in a single- or double-effect evaporator. Water is condensed from the vapor to leave concentrated (95 percent) SO_2 gas. The Na_2SO_3 solids are redissolved and recycled to the absorber. The concentrated SO_2 is converted to sulfur by reaction with methane[44] or to sulfuric acid by reaction with air. When the process is used to treat tail gas from sulfuric acid or sulfur recovery plants, the concentrated SO_2 is recycled to the front end of the plant.

With most waste gases, prescrubbing is necessary to remove residual particulates, HCl, and SO_3. In the absorber, oxygen present in the flue gas converts some sulfite to sulfate. Typically this sulfate must be removed from the absorbent solution by the selective crystallization of sodium sulfate. The evaporative crystallizer must be operated below 100°C to prevent excessive disproportionation of bisulfite to sulfate and thiosulfate.

The Wellman-Lord process was first tested in 1970 on tail gas from a sulfuric acid plant in Paulsboro, New Jersey. There followed several applications on tail gas from sulfur recovery plants in the United States and Japan and on flue gas from oil-fired boilers in Japan. The first application to a coal-fired boiler was made in 1977 in the United States.

Magnesium Oxide Process. The MgO process has been tested in the United States, but no commercial units are currently operating.[20] As shown in Figure 4-4, SO_2 is absorbed in a slurry of $MgSO_3$. Makeup MgO is added to the

Figure 4-3. Wellman-Lord process.

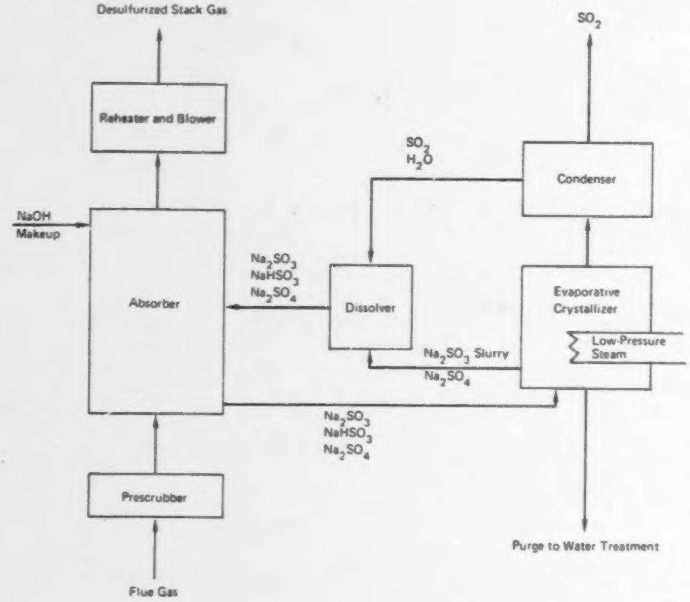
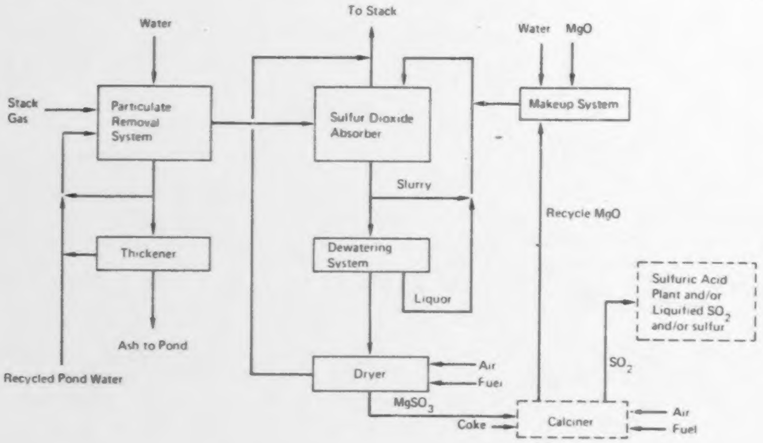

Desulfurized Stack Gas
Reheater and Blower
NaOH Makeup
Absorber
Prescrubber
Flue Gas
Na_2SO_3
$NaHSO_3$
Na_2SO_4
Dissolver
SO_2
H_2O
SO_2
Condenser
Evaporative Crystallizer
Low-Pressure Steam
Na_2SO_3 Slurry
Na_2SO_4
Na_2SO_3
$NaHSO_3$
Na_2SO_4
Purge to Water Treatment

Figure 4-4. Magnesium oxide scrubbing.


Water
Stack Gas
Particulate Removal System
Thickener
Ash to Pond
Recycled Pond Water
To Stack
Sulfur Dioxide Absorber
Slurry
Dewatering System
Liquor
Dryer
Air
Fuel
$MgSO_3$
Coke
Calciner
Air
Fuel
Water
MgO
Makeup System
Recycle MgO
SO_2
Sulfuric Acid Plant and/or Liquified SO_2 and/or sulfur

recirculating slurry to produce $MgSO_3$; $MgSO_3 \cdot 3H_2O$ or $MgSO_3 \cdot 6H_2O$ solids are separated by a centrifuge and dried to remove entrained water and water of hydration. The anhydrous $MgSO_3$ is shipped to a sulfuric acid plant where it is converted to MgO and SO_2 at 815-1090°C by direct-fired heating in a rotary or fluidized bed calciner. The gas stream is cleaned and used as a source of SO_2 for sulfuric acid production. The MgO solids are shipped back to the scrubber site.

The process was tested on stack gas from an oil-fired boiler at Boston Edison in 1972-1974.[21] It has also been tested on stack gas from coal-fired boilers at Potomac Electric Power Company[22] and Philadelphia Electric Company.[23] In all three cases there were problems with solids handling and with coordination between the scrubbing and regeneration facilities. However, the process appears to be basically sound and is awaiting application in an economically attractive situation.

Dry Processes

Dry processes avoid saturating the flue gas. A number of such systems have been partially developed, as shown in Table 4-2. None of them has been used commercially in the United States, however, primarily because of the capital cost required for handling fixed or moving beds of solids.

Carbon absorption has been extensively tested on large units in Japan, Germany, and the United States. Catalytic oxidation was tested in a 100-MW unit in the United States. CuO absorption with H_2 regeneration has been used in Japan and tested briefly in the United States. Both CuO adsorption and several activated-carbon processes have the potential for removing some nitrogen oxides along with the sulfur oxides, if ammonia is added.

WET SCRUBBING EQUIPMENT

Wet scrubbing using lime or limestone is the predominant technology used in the United States and Japan, which are the two nations using most of the postcombustion sulfur oxides cleanup systems. The basic element of these wet-process systems is the absorber in which the flue gas is placed in contact with the scrubbing medium.

Although FGD systems come in many sizes, shapes, and configurations, primary consideration has been given to four general types: venturi systems with or without spray towers (Figure 4-5), turbulent contact absorbers (Figure 4-6), packed or tray-type absorbers (Figure 4-7), and horizontal systems (Figure 4-8).

Table 4-2. Dry processes.

Generic Name	Process Operation	Product
Carbon adsorption		
Water wash[45]	Absorption at 150° C; regenerate to dilute H_2SO_4; $CaCO_3$ neutralization	$CaSO_4$
Thermal[45-47]	Absorption at 150° C; desorption at 300-600° C	10-20% SO_2
Reducing gas[48]	Absorption at 150° C; desorption at 300-600° C with H_2	S
CuO adsorption[49,50]	Adsorption on CuO at 350° C; desorption by H_2 at 350° C	30-90% SO_2
Catalytic oxidation[51]	Catalytic oxidation at 450° C; absorption in H_2SO_4 at 150° C	85% H_2SO_4
Activated Na_2CO_3[52]	Absorption by nahcolite or trona at 150° C	Na_2SO_4
Dry limestone[53]	Adsorption by $CaCO_3$ or dolomite at 600-1000° C	$CaSO_4$
Molten carbonate[54]	Absorption in molten Li, Na, K_2CO_3 at 400° C; reaction with coke at 700° C; carbonation	H_2S
MnO_2[45]	Absorption by MnO_2 at 150° C; aqueous oxidation; NH_3 neutralization	$(NH_4)_2SO_4$

Figure 4-5. Schematic of venturi system and spray tower.

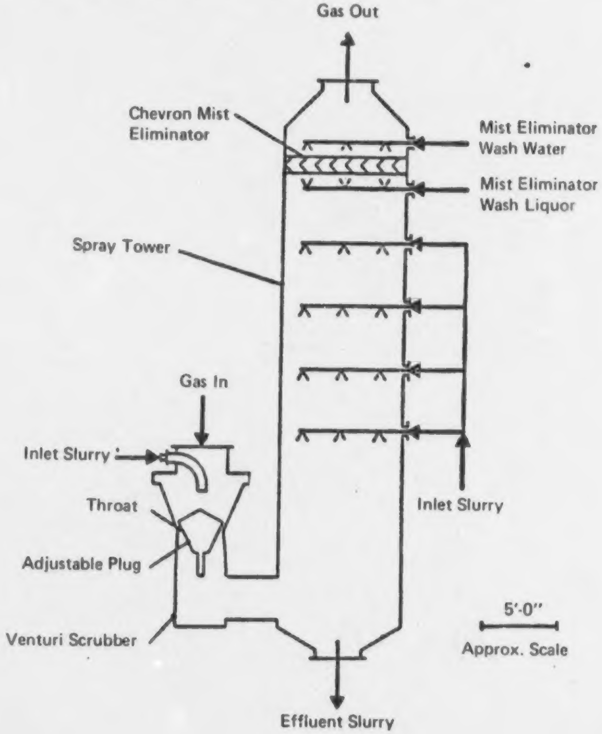

Gas Out
Chevron Mist Eliminator
Mist Eliminator Wash Water
Mist Eliminator Wash Liquor
Spray Tower
Gas In
Inlet Slurry
Throat
Adjustable Plug
Venturi Scrubber
Inlet Slurry
5'-0"
Approx. Scale
Effluent Slurry

Figure 4-6. Schematic of three-bed turbulent contact absorber system.

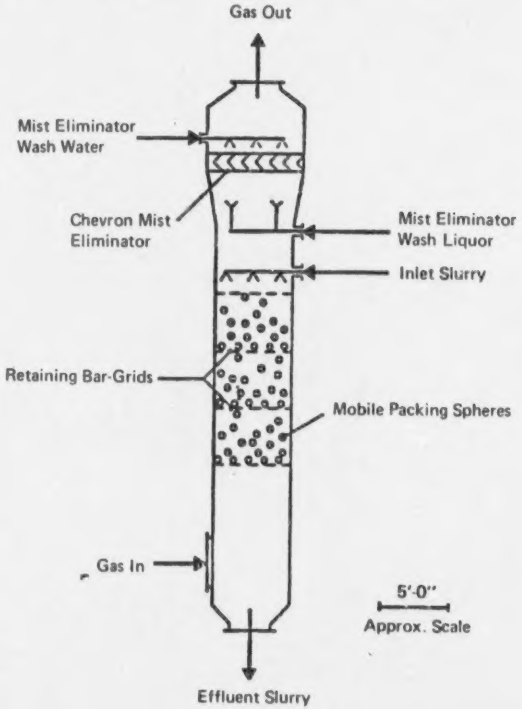

Gas Out
Mist Eliminator Wash Water
Chevron Mist Eliminator
Mist Eliminator Wash Liquor
Inlet Slurry
Retaining Bar-Grids
Mobile Packing Spheres
Gas In
5'-0"
Approx. Scale
Effluent Slurry

Figure 4-7. Packed absorber.

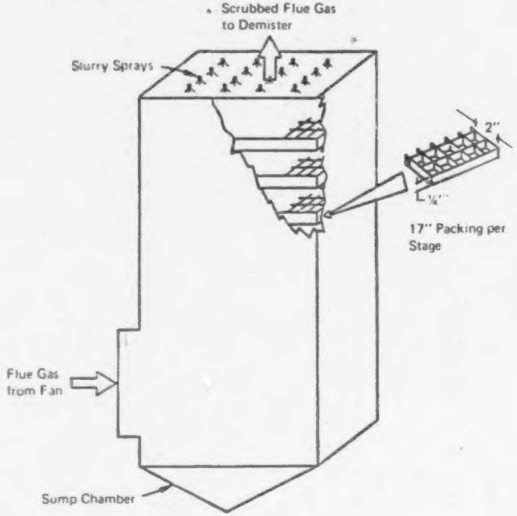
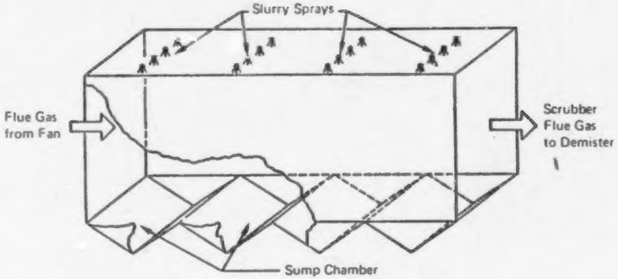

Scrubbed Flue Gas to Demister
Slurry Sprays
2"
¼"
17" Packing per Stage
Flue Gas from Fan
Sump Chamber

Figure 4-8. Horizontal scrubber.


Slurry Sprays
Flue Gas from Fan
Scrubber Flue Gas to Demister
Sump Chamber

The first three of these units use a vertical configuration and vertical countercurrent flows of flue gases and scrubbing media. The fourth involves horizontal flow of the flue gases and vertical flow of the scrubbing medium.

Each general type of FGD process has its advantages, but operational difficulties have also been encountered with each. Three major equipment-related problem areas have been identified[24] in lime and limestone scrubbing; these are mist elimination, materials of construction, and the need for a better understanding of process chemistry and its relation to system design. This is discussed in more detail in Chapter 6.

COMMERCIAL STATUS OF FLUE-GAS DESULFURIZATION TECHNOLOGY

The status of FGD systems for the U.S. electric utility industry as of November 1978 is given in Table 4-3.[26] There are 46 operational units, and an additional 98 units are planned or under construction. Table 4-4 gives the distribution of FGD systems on utility boilers by chemical processes. Lime or limestone slurry scrubbing processes account for 88 percent of the total. Wellman-Lord, MgO, and dual alkali processes each account for 2 to 4 percent of the total.

Table 4-5 gives the FGD plants operational in Japan as of December 1978.[27] The total installed FGD capacity in Japan is equal to 26,000 MW, or 50 percent higher than that in the United States, although about half of this capacity is on nonutility boilers.

A description of the individual FGD units in operation in the United States and a discussion of their efficiency and reliability are given in Chapter 6. Although existing FGD installations are dominated by lime and limestone systems, others are under development and expected to become commercially available in the future. The reliability, efficiency, and operating problems of these new processes are still not known.

CROSS-MEDIA IMPACTS OF FLUE-GAS CLEANING PROCESSES[56]

Air pollution control technology produces cross-media environmental impacts from the solids in the coal, from the FGD wastes, from the increased air pollution created by boiler derating, and from changes in the nature of the pollutants emitted.

The Resource Conservation and Recovery Act of 1976 (RCRA) is designed to regulate all solid-waste disposal to

Table 4-3. Number and capacity of U.S. utility FGD systems.

Status	Number of Units	Capacity (MW)
Operational	46	16,054
Under construction	43	17,297
Planned		
Contract awarded	20	10,690
Letter of intent signed	3	1,960
Requesting or evaluating bids	5	3,100
Considering FGD	27	13,406
Total	144	62,507

Table 4-4. Distribution of FGD systems by chemical process.

	FGD Capacity (MW)		
Process	Operational	Under Construction	Planned
Limestone[a]	8,734	8,687	10,848
Lime[b,c]	6,070	6,029	3,482
Lime/limestone	20	330	330
Sodium carbonate	375	509	0
MgO	120	0	1,326
Wellman-Lord	735	180	940
Dual alkali	0	1,102	0
Aqueous carbonate[d]	0	400	100
Citrate[e]	0	60	0
Total[f]	16,054	17,297	17,026

[a] Includes alkaline-fly-ash-limestone and limestone-slurry configurations.

[b] Includes alkaline-fly-ash-lime and lime-slurry configurations.

[c] Includes nonregenerable dry-collection and nonregenerable wet scrubbing configurations.

[d] Includes nonregenerable dry collection and regenerable configurations.

[e] This system is being installed at St. Joseph Minerals' G. F. Wheaton Plant and is listed as a utility FGD system because the plant is connected by a 25-MW interchange to the Duquesne Light Company.

[f] Because the processes for all planned systems are not known, the totals in this table are less than those in Table 4-3.

Table 4-5. FGD plants in Japan operational at end of 1978.

Plant Constructor	Wet Lime/Limestone		Indirect Lime/Limestone		H_2SO_4S, $(NH_4)_2SO_4$		Na_2SO_3, Na_2SO_4		Total	
	No. of Plants	Nm^3/hour[a]	No. of Plants	Nm^3/hour[a]	No. of Plants	Nm^3/hour[a]	No. of Plants	Nm^3/hour[a]	No. of Plants	Nm^3/hour[a]
Mitsubishi Heavy Industries (MHI)	34	19 020					3	292	37	19 312
Ishikawa[illegible] Heavy Industries (IHI)	17	4 [illegible]					79	4 351	96	8 796
Hitachi, Ltd.	13	6 [illegible]			2	590	15	603	30	8 133
Mitsubishi Kakoki (MKK)	2	256			13	6 478	41	913	56	7 647
Kawasaki Heavy Industries	4	[illegible]	[illegible]	5 450			7	256	17	6 462
Tsukishima Kikai (TSK)	1		[illegible]	398	1	88	40	4 042	46	4 528
Chiyoda Chemical Engineering and Construction			15	4 585					15	4 585
Oji Koei							57	4 260	57	4 280
Sumitomo Metal-Fujikasui	7	3 954					6	270	13	4 224
Kurabo Engineering			5	603	1	18	106	3 751	112	4 372
Mitsui Miike-Chemico	4	2 744			1	500			5	3 244
Ebara Manufacturing			11	1 914			10	1 167	21	3 081
Kobe Steel	6	2 475							6	2 475
Nippon Kokan (NKK)	3	245	1	150	2	1 990	6	62	12	2 447
Kureha Chemical							8	1 431	8	1 431
Showa Denko							5	1 372	5	1 372
Cadelius							8	1 291	8	1 291
Sumitomo (SCEO)-Wellman					6	1 288			6	1 288
Nippon Steel	2	1 200							2	1 200
Mitsui Metal Engineering	4	1 006			2	130			6	1 136
Dowa Engineering			8	666					8	666
JGC	1	330			1	125			2	455
Ube Industries					2	220			2	220
Niigata Engineering			1	185					1	185
Mitsui Engineering							1	160	1	160
Total	98	43 371	51	13 951	31	11 427	392	24 241	572	93 150

[a] Nm^3/hour = normal cubic meters per hour; 1000 Nm^3/hour = 0.27 mw equivalent gas capacity.
[b] Babcock Hitachi.
[c] Wellman NKK.

prevent pollution and minimize health risks. It is intended to abolish open dumps, require permits for all hazardous-waste deposits, and impose requirements for solid-waste disposal. Hazardous wastes are those that contribute to illness or pose hazards to either health or the environment.

The basic regulations to implement this act have not yet been adopted, but the facility generating the waste will be required to determine, by test, whether the waste is hazardous and, if it is, must keep strict records and be accountable for the materials for 20 years.

Electric utility fly ash and solid wastes generated by flue-gas desulfurization are to be classified initially as special large-volume hazardous wastes, which will have interim facility standards. Costs of disposal under these circumstances are estimated to be about 88 cents per million Btu, compared with 31 cents per million Btu using current utility prices.[55] No final decision has been made on how coal ash or FGD solid wastes will eventually be classified, but if for some plants they are classified as hazardous wastes, disposal costs could be increased by a factor of 10 or more.

Technology and Production of Wastes

Coal-fired utility and industrial boilers generate two types of coal ash--fly ash and bottom ash. Depending on the way in which the boiler is fired and on the fusion temperature of the ash, as much as 70 to 95 percent of the ash may be carried out of the boiler with the flue gas as fly ash, the remainder being removed as bottom ash. Total coal-ash production by electric utilities is expected to reach 72 million tons/year (including over 45 million tons/year of fly ash) by 1980,[25] and industrial boilers are expected to increase this amount by 10 to 15 percent.

As noted above, nonrecovery throwaway FGD systems comprise 90 percent of the utility FGD commitments, and most of them generate a solid waste; the remainder generate a liquid waste.

The quantities of coal ash and FGD wastes produced depend primarily on the ash and sulfur contents of the coal, the applicable air emission regulations, the type of FGD system employed, and the operating conditions of both the FGD system and the boiler. To meet NSPS of 1.2 pounds of SO_2 per million Btu, a typical utility boiler operating at a 70 percent load factor can produce annually between 100 and 500 short tons of dry, ash-free solid waste per megawatt of capacity from the scrubber and from one to two times as much ash. The total quantity of coal ash and FGD waste is expected to be 80 million tons or more (dry basis) by 1980.

Characterization of Wastes

Coal Ash. The chemical composition of coal ash (bottom ash and fly ash) varies widely in concentrations of both major and minor constituents. The principal factor affecting the variation of the composition is the variability in the mineralogy of the coal; however, differences in composition can also exist between fly ash and bottom ash generated from the same coal. Regardless of the type of ash (either fly ash or bottom ash), more than 80 percent of its total weight is usually made up of silica, alumina, iron oxide, and calcium salts. As much as 20 percent of fly ash can be water soluble, so the potential exists for release of contaminants through leaching.

Coal also contains a number of trace elements. Concentrations of trace elements tend to be highest in eastern coals (Appalachian) and lowest in western coals.[29] On the average, only four elements--arsenic, boron, chlorine, and selenium--are present in coals in concentrations much above the average concentration in the earth's crust.

Although the major constituents of bottom ash and fly ash are generally similar, there may be a substantial enrichment of a number of trace elements in the fly ash (antimony, selenium, and lead) as compared with bottom ash.[30] A few of the elements originally present in coal (notably sulfur, mercury, and chlorine) enter the gas stream in some types of combustion equipment and leave the boiler as gaseous species that are not collected downstream in dry-ash collection equipment; however, these may be collected in wet FGD systems used for particulate and/or sulfur oxides control.

FGD Wastes. FGD waste characteristics, particularly the chemical composition, may vary over extremely wide ranges. The principal constituents in solid wastes from most nonrecovery FGD systems are calcium-sulfur salts (present as calcium sulfite and/or calcium sulfate), although significant quantities of calcium carbonate, unreacted lime, inerts (inorganic compounds not oxidized or otherwise reacted), and/or fly ash can also be present. The ratio of calcium sulfite and calcium sulfate (the latter present as $CaSO_4 \cdot 1/2\ H_2O$ or as gypsum, $CaSO_4 \cdot 2\ H_2O$) will depend on the extent to which oxidation occurs within the system.

Combined fly ash and FGD wastes are referred to by the generic term, flue-gas cleaning (FGC) wastes. Fly ash will be a principal constituent of FGC wastes only if the scrubber serves as a particulate control device in addition to removing sulfur oxides or if separately collected fly ash is admixed with the FGD wastes. The amounts of inerts and unreacted raw materials (lime and/or limestone) will depend on system stoichiometry.

FGD waste solids carry with them occluded or free liquor (10 to 90 percent, depending on the amount of dewatering) that contains a wide variety of dissolved substances ranging from trace amounts of various heavy metals to substantial quantities of commonly occurring species such as sodium, calcium, magnesium, chloride, and sulfate. In general, the concentrations of different chemical species in solution will be dictated either by equilibrium solubilities (as in the case of most calcium salts) or, for the most highly soluble species, by the rate at which they enter the FGD system.

The levels of trace elements in FGC wastes depend primarily on four factors:

- Levels of trace constituents in the coal relative to its sulfur content;
- Presence and levels of trace elements in FGD-process additives and makeup water;
- Amount of ash, if any, collected with or admixed with the FGD wastes; and
- Efficiency of the scrubber system in capturing volatile trace constituents.

Data from the analysis of wastes produced by direct lime and limestone wet-scrubbing systems indicate that most of the trace elements in FGC wastes result from the presence of fly ash.[30] Tables 4-6 and 4-7 give the ranges of concentrations of trace elements and soluble chemical species measured in samples of ash-rich wastes and waste liquors.

Waste liquors from once-through sodium scrubbing usually consist primarily of sodium sulfate, chloride, and sulfite (if not oxidized to sulfate as a part of the system operation) at total dissolved solids levels up to 30 percent by weight. The waste may also contain appreciable quantities of fly ash when the system is used for simultaneous sulfur oxides and particulate control.

Disposal Options and Regulatory Factors

At present, coal ash and FGC wastes are disposed of exclusively on land. The three principal disposal methods are wet ponding, dry impoundment, and mine disposal.

Table 4-6. Typical concentrations of trace elements in FGC wastes.

Element	Concentration Ranges (ppm)	Median Concentration (ppm)[a]	Number of Observations	Range of Trace Element Measured in Coal (ppm)
Arsenic	3.4-63	33	9	3-60
Beryllium	0.62-11	3.2	8	0.08-20
Cadmium	0.7-350	4.0	9	—
Chromium	3.5-34	16	8	2.5-100
Copper	1.5-47	14	9	1-100
Lead	1.0-55	14	9	3-35
Manganese	11-120	63	5	—
Mercury	0.02-6.0	1	9	0.01-30
Nickel	6.7-27	17	5	—
Selenium	0.2-19	7	9	0.5-30
Zinc	9.8-118	57	5	0.9-600

[a] Values as reported.

Source: R. R. Lunt et al.[31]

Wet Ponding. Wet ponding usually involves discharge of FGC waste slurries (either directly from the scrubber or after thickening) to a diked containment area where waste solids settle out and supernatant liquor is discharged or recycled to the FGC system. This is a common method used for temporary storage or ultimate disposal of FGC wastes. Evaporation-settling ponds are also used for disposal of liquid wastes from once-through sodium scrubbing systems. In some cases, solid wastes are pretreated before discharge by addition of chemicals to stabilize the wastes, resulting in a hardened material of reduced permeability.

Dry Impoundment. In dry impoundment, moist dewatered solid wastes, usually admixed with dry fly ash, are deposited on a graded site and compacted. Dry impoundment is being used increasingly, frequently in conjunction with stabilization processes, for minimizing the environmental impact of waste disposal. Stabilization, as now practiced, usually involves admixing filtered or centrifuged wastes with dry fly ash and a small amount of lime, which improves the ultimate structural stability and reduces the leaching potential (through reduced permeability and possibly some insolubilization of soluble species).

Disposal in Mines. Mine disposal is potentially attractive because the same transport system that brings the

Table 4-7. Typical concentrations of chemical species in FGD sludge liquors and elutriates.

	Eastern Coals			Western Coals		
Species	Range in Liquor (ppm)	Median (ppm)	Number of Observations	Range in Liquor (ppm)	Median (ppm)	Number of Observations
Antimony	0.46-1.6	1.2	4	0.09-0.22	0.16	2
Arsenic	<0.004-1.8	0.020	15	<0.004-0.2	0.009	7
Beryllium	<0.0005-0.05	0.014	16	0.0006-0.14	0.013	7
Boron	41	41	1	8.0	8.0	1
Cadmium	0.004-0.1	0.023	11	0.011-0.044	0.032	7
Calcium	470-2,600	700	15	240-~45,000	720	6
Chromium	0.001-0.5	0.020	15	0.024-0.4	0.08	7
Cobalt	<0.002-0.1	0.35	3	0.1-0.17	0.14	2
Copper	0.002-0.4	0.015	15	0.002-0.6	0.20	7
Iron	0.02-0.1	0.026	5	0.42-8.1	4.3	2
Lead	0.002-0.55	0.12	15	0.0014-0.37	0.016	7
Manganese	<0.01-9.0	0.17	8	0.007-2.5	0.74	6
Mercury	0.0009-0.07	0.001	10	<0.01-0.07	<0.01	7
Molybdenum	5.3	5.3	1	0.91	0.91	1
Nickel	0.03-0.91	0.13	11	0.005-1.5	0.09	6
Selenium	<0.005-2.7	0.11	14	<0.001-2.2	0.14	7
Sodium	36-20,000[a]	118	6	1,650-~9,000[a]	—	2
Zinc	0.01-27	0.046	15	0.028-0.88	0.18	7
Chloride	470-5,000	2,300	9	1,700-43,000[b]	—	2
Fluoride	1.4-70	3.2	9	0.7-3.0	1.5	3
Sulfate	720-30,000[a]	2,100	13	2,100-18,500[a]	3,700	7
TDS	2,500-70,000[a]	7,000	—	5,000-95,000[b]	12,000	3
pH	7.1-12.8	—	—	2.8-10.2	—	3

[a] Levels of soluble sodium salts in dual-alkali sludge (filter cake) depend strongly on the degree of cake wash. The highest levels shown reflect single measurements on an unwashed dual-alkali filter cake.

[b] Levels of soluble chloride components in sludges are dependent on the chloride-to-sulfur ratio in the coal. The highest levels shown are single measurements for a Western limestone scrubbing system operating in a closed loop using cooling-tower blowdown for process makeup water.

Source: R. R. Lunt et al.[31]

coal to the power plant can be used to transport the wastes to the mine. It also eliminates the need for large tracts of new land for disposal. Placement of FGC wastes in mines may also be beneficial in that it fills voids left from coal extraction. In general, active surface mines are considered to be the most attractive sites.[31] In surface mines, dewatered wastes mixed with fly ash or otherwise stabilized could be placed either in the pit bottom before replacement of overburden or in the overburden spoil banks before reclamation.

Ocean Disposal. In areas where no mines are available and disposal sites for land impoundments are scarce, ocean disposal may be a technically feasible alternative. It is now being jointly studied by EPA, DOE, and EPRI.

Regulatory Factors. The disposal of FGC waste and coal ash is and will continue to be subject to federal and state regulations. The key legal framework concerning environmental impact on land is the Resource Conservation and Recovery Act, which provides sufficient statutory basis for preventing significant adverse health and environmental impacts. However, much of the development and implementation of specific regulations lies ahead.

Cross-Media Impact Issues

Potential environmental impacts of FGC systems are dependent on a multiplicity of factors and their complex interrelationships, so that impacts tend to be site specific and cannot be easily generalized. Therefore, an examination of cross-media impacts outside the context of a specific application of FGC technology can only address the potential environmental impact issues. The following discussion identifies the principal cross-media environmental issues related to air, water, and land impacts.

Air Quality Impacts. The FGC system is a control measure designed to reduce facility emissions, but it can produce other air quality effects. These may be divided into three categories: changes in facility emission levels, changes in pollutant distribution, and fugitive emissions from waste disposal.

The application of commercially available FGD technology for sulfur oxides control causes a boiler derating of 3 to 5 percent of its generating capacity because of the consumption of power and steam (for reheat) by the FGD system. A resultant equivalent increase in emissions of nitrogen oxides would be expected because of the extra coal combustion required to provide this heat and energy. The impact on the amount of particulate emissions depends on plant design, as does the amount of nitrogen oxides, since some of the oxides are removed by the scrubber.

Another complicated consideration is the extent to which FGD systems might influence the formation and emission of fine particulate sulfates. The overall reduction in sulfur oxides emissions resulting from FGD system operation would be expected to reduce the overall potential for sulfate formation by SO_2 oxidation. However, a wet scrubber system will also change the humidity in the immediate vicinity and might change the levels of trace metals in the emission plume. Both of these factors have been considered important in the oxidation of SO_2 to sulfate, but data to evaluate the significance, if any, of these combined changes are not currently available.

Data to permit evaluation of the significance of changes in pollutant distribution resulting from FGC system operations are lacking. However, FGD system operations might be expected to influence the extent of both wet and dry deposition of various pollutants in the vicinity of a power plant.

Water-Related Impacts. Water-related impacts can be broken down into two categories, point source and nonpoint source.

Point Sources. Potential point-source impacts stem from the discharge of noncontact cooling and seal water from FGC systems as well as from the discharge of any process liquor. The primary effect of noncontact water on water quality is in terms of sensible heat and oil and grease. The quantity of noncontact cooling water is relatively small in comparison with other plant waste streams, and these discharges would normally be treated along with other plant wastewater and/or be recycled for reuse either in the FGC system or elsewhere in the plant.

Direct discharges of process liquors from solid-waste disposal operations could occur in the form of supernatants from wet impoundments. The other types of point-source impact for process water discharge from FGD systems are from liquid-waste-producing nonrecovery systems and from prescrubber blowdown from recovery systems. The only widely applied FGD system that produces liquid waste involves once-through sodium scrubbing using some form of soda ash, caustic, or impure brine. The waste liquor, which usually amounts to about 0.5 to 1.5 gpm per megawatt of capacity for high-sulfur coal systems, can be discharged to evaporation ponds or wastewater treatment systems. If fly ash is present, the wastes are usually discharged to a settling pond before subsequent treatment or disposal. For large-scale systems, evaporation ponds are generally used.

Prescrubber blowdown from recovery systems is usually small in volume, generally 0.1 to 0.2 gpm per megawatt of capacity. However, the blowdown can be quite acidic ($pH < 2.0$) and can contain up to a few thousand parts per million of dissolved solids, as much as 5 to 10 percent suspended solids (fly ash and calcium-sulfur salts), and numerous trace metals (as cited earlier). The waste stream would normally be neutralized and solids removed as a part of the FGC system operation. The clarified, neutralized liquor would then be either mixed with other plant effluents or further treated.

Nonpoint Sources. The solid wastes generated by FGC systems have the potential to affect water quality, especially in conjunction with their disposal. The potential impact mechanisms of principal interest for all land-based disposal options are leachate and runoff.

The wet-ponding disposal method has maximum potential for leaching of constituents because of the constant intimate contact with water. The extent of leaching depends on the permeability of the wastes and underlying soil, surface runoff, and other factors. To reduce the potential for transport of contaminants into groundwater, ponds can be constructed in areas with low-permeability soils or can be lined with clay.

Major soluble contaminants of importance in FGC wastes are the sulfates, sulfites, and chlorides. For certain dry-sorbent systems using sodium alkalis, overall waste solubility (sodium availability, in particular) could have important adverse impact potential in disposal environments close to usable freshwater resources. Sodium may also be important in leachate or runoff from impoundments of wastes from sodium carbonate scrubbing systems. Stabilization processes have been applied to the wastes from wet FGC systems to achieve reductions in permeability and the rates of leaching of major species. However, runoff (and, to a degree, leachate) quality may still require attention on a site-specific basis in the disposal of stabilized FGC wastes, especially during their placement.

FGC wastes can exhibit relatively high trace-metal levels. From the standpoint of water quality impacts, the presence and availability of metals such as As, B, Se, Cd, Ni, and Zn can be of importance in the leachate and/or runoff from FGC waste-disposal areas or in the ocean disposal of such wastes.

Land-Related Impacts. Geologic and soil-related impacts are associated with the stability of the FGC wastes that are disposed of. Physical instability is a potential problem for all FGC wastes, including stabilized wastes in all types of land disposal. Important considerations are design factors affecting disposal site configurations, stability of underlying materials, erosion potential, and climatic considerations including freeze-thaw conditions.

Instability problems can be ameliorated by compaction methods using dry impoundments and disposal in mines.

Impact on land used for waste disposal is another important environmental factor. A typical 1000-MW power plant requires over a 30-year plant life an estimated 160 to 280 hectares (400 to 700 acres) for disposal of ash and FGD wastes when piled to a height of 20 to 30 feet. This refers to the excavated area proper; actual land commitment, including access roads and buffer zones, may be much larger. However, in terms of national or regional land use, the total land requirements for FGC waste disposal are not significant. If all the coal ash and FGD wastes are deposited by dry-disposal methods to a depth of 10 m (32.8 feet), the total cumulative land commitment (excavated area only) for these wastes will be 5580 hectares (13,750 acres) by 1985 and 19,600 hectares (48,150 acres) by 2000. However, the land required in a given locality could require modifications of land use planning and practices on a site-specific basis.

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This article, from a Noyes data report on flue gas technology, provides a description of various FGD systems, including a detailed analysis of lime/limestone FGD, the systems most in use this past decade. Tables 1.1-1.3 give a good summary of the various processes and their commercial status. Table 1.4 lists the important facts concerning FGD on boiler facilities, including removal efficiency and system availability (both important in determining cost effectiveness of a process).

INTRODUCTION

The information in the following three sections is based on *Flue Gas Desulfurization Pilot Study—Phase I—Survey of Major Installations—Summary of Survey Reports on Flue Gas Desulfurization Processes,* January 1979, NTIS PB-295 001, prepared by the Flue Gas Desulfurization Study Group of the NATO Committee on the Challenges of Modern Society (NATO-CCMS Study), F. Princiotta of the U.S. Environmental Protection Agency, Chairman.

BACKGROUND

The first commercial application of flue gas desulfurization (FGD) to power plant sulfur oxide control was in the United Kingdom in the early 1930s. The Battersea A Power Plant (228 MW) of the London Power Company, London, UK, began flue-gas washing in 1933. The process utilized wet scrubbing with Thames River water providing most of the alkaline absorbent. The spent absorbent was discharged back into the Thames after settling and oxidation. The FGD system operated successfully at up to 95% SO_2 removal efficiency until the Battersea A Power Plant closed down in 1975. A similar FGD system operated on the Battersea B Power Plant (245 MW) between 1949 and 1969, when FGD operation was temporarily suspended because of adverse effects on the Thames water quality.

The ICI Howden process, also developed in England, was developed to avoid discharging scrubber effluent into the Thames. A solid sludge was produced and barged to sea for dumping. This process was applied to the Swansea Power Plant in 1935 and the Fulham Power Plant in 1937. These systems operated successfully until early World War II when they were shut down.

The next FGD unit was installed at the Electrolytic Zinc Company in Tasmania in 1949. Tidal water was used there as the absorbent for SO_2 from smelter gas.

In 1952 the first unit of the new oil-fired Bankside Power Station in London, UK, was commissioned. This FGD system is an improved version of the Battersea system, using water from the Thames. This system is still operating at up to 98% removal efficiency and with a present capacity of 240 MW.

The 1950s and 1960s were a time of laboratory and pilot plant investigations of new processes. During the 1950s the Tennessee Valley Authority (USA) investigated lime/limestone systems, both dry and wet, and dilute acid processes; in Germany the first major carbon adsorption processes were developed. During the 1960s the magnesium oxide, copper oxide, and sulfite scrubbing processes were investigated among others.

Lime/limestone processes were installed in 1964 on an iron ore sintering plant in Russia and on a large sulfuric acid plant in Japan in 1966.

In 1966 Combustion Engineering developed a dry limestone injection process, which was installed at five boilers in the United States by 1972. Because of major problems associated with dry limestone injection including plugging (especially of the boiler tubes), low sulfur dioxide removal, and reduced particulate collection in the electrostatic precipitators, these systems proved inadequate. The five installations are now either closed down or converted to other control systems.

Japan has at present the largest installed capacity of FGD systems. Most of these systems were built between 1975 and 1977. Systems based on lime/limestone predominate. As of December 1977, the United States had 27 operating utility units treating about 35 x 10^6 Nm^3/h (normal conditions are 0°C and 1 bar), 34 units under construction to treat 50 x 10^6 Nm^3/h, and 20 units planned to treat 34 x 10^6 Nm^3/h; in addition 25 industrial boilers were operational or under construction with a capacity of about 5 x 10^6 Nm^3/h. In the Federal Republic of Germany, approximately 775,000 Nm^3/h of flue gas are being treated by lime and carbon adsorption processes, and in Norway 155,000 Nm^3/h are treated by seawater scrubbers on oil-fired boilers.

PROCESS CATEGORIES

FGD processes can most conveniently and usefully be categorized by the manner in which the sulfur compounds removed from the flue gases are eventually produced for disposal. In this way three main categories result:

(1) Throwaway processes, in which the eventual product is disposed of entirely as waste. Disposal can include landfill, ponding, discharge to water course or ocean, or discharge to a worked-out mine.

The processes in this category involve wet scrubbing of the flue gases for absorption, followed by various methods for neutralizing the acidity, separating the sulfur compounds from the scrubbing liquor, and usually recycling at least part of the scrubbing liquor.

(2) Gypsum processes, which are designed to produce gypsum of sufficient quality either for use as an alternative to natural gypsum or as a well-defined waste product with good disposal characteristics.

As with the throwaway processes, this category involves wet scrubbing for absorption followed by various methods of neutralizing the lime or limestone and recovering the sulfur compound. An oxidation step is included to insure recovery of the sulfur compounds in the form of gypsum.

(3) Regenerative processes, which are designed specifically to regenerate the primary reactants and concentrate the sulfur dioxide that has been removed from the flue gases. Further chemical processing can then convert the concentrated SO_2 into sulfuric acid or elemental sulfur, or physical processing into liquefied sulfur dioxide. The surveyed processes in this catagory contain both wet scrubbing and dry adsorption processes.

Tables 1.1, 1.2, and 1.3 present a summarized description of processes in each of these 3 categories.

STATUS OF OPERATING FGD SYSTEMS

There are now 144 known FGD systems operating on fossil-fueled combustion sources. Of these, 74 are operating on power utility boilers, representing about 70 x 10^6 Nm^3/h of flue gas capacity from about 20 GW generation. The remainder are operating on industrial combustion sources, principally boiler plants, but also on iron ore sinter plants and petroleum refinery plants.

In the NATO-CCMS Process Status Reports, 35 FGD systems were surveyed. Although it was originally desired to include only the larger, commercially available installations with adequate operating experience, it was necessary to include data on some smaller scale operations and some that had been installed originally for demonstration purposes in order to provide sufficient comparability between the various processes.

Table 1.4 summarizes basic technical data on the FGD systems surveyed in the NATO-CCMS study and presents information on fuel burned and system operation.

Throwaway Processes: This category includes the three processes that produce a calcium sulfite/sulfate sludge and also the seawater scrubbing process. The sludge processes are becoming the most widely used FGD systems; 30 systems are in operation and 53 are planned or under construction. They have been used successfully on both coal- and oil-fired plants with a wide range of fuel sulfur contents and are reported to have high SO_2 removal efficiencies, whereas plant availabilities are variable.

The double alkali process is a further development designed to overcome the scaling, plugging, and erosion problems that have been generally associated with lime or limestone systems. Double alkali systems are presently used mainly with small to medium sized boilers where the extra process equipment can be offset by lower maintenance costs.

Gypsum Processes: These processes are designed to produce a quality of gypsum that may be used in place of natural gypsum in such markets as plaster or plaster wallboard, or as a setting retarder in cement manufacture. It is expected that if sufficient markets are not available for gypsum, it will be disposed of as a solid waste. Gypsum has better settling characteristics than sludges containing calcium sulfite from the throwaway processes.

Table 1.1: Summary Description of FGD Processes—Throwaway Category

FGD Process (year of first use)	Primary Reagent	Product for Disposal	Operating Principles	Present Status
Limestone/sludge (1933)	$CaCO_3$	$CaSO_3/CaSO_4$ wet sludge	Flue gases scrubbed with limestone slurry. SO_2 reacts to form insoluble calcium compounds, which are separated and partly dewatered for disposal.	Commercial operation (coal and oil), principally in U.S.
Lime/sludge (1972)	CaO	$CaSO_3/CaSO_4$ wet sludge	Flue gases scrubbed with slurry of hydrated lime. SO_2 reacts to form insoluble calcium compounds, which are separated and partly dewatered for disposal.	Commercial operation (coal and oil), principally in U.S.
Double alkali/sludge (1974)	NaOH or Na_2CO_3 and CaO or $CaCO_3$	$CaSO_3/CaSO_4$ wet sludge	Flue gases scrubbed with alkaline sodium based compounds in solution. Used solution regenerated with lime or limestone in separate reactor to form insoluble calcium compounds, which are separated and partly dewatered for disposal.	Recent commercial operation on industrial sized coal-fired boilers in U.S.
Seawater scrubbing (1975)	Seawater	Dilute sulfate solution	Flue gases scrubbed with seawater on a once-through system, diluted, oxidized and neutralized, and discharged back to sea.	Commercial operation on small oil-fired boilers in Norway.

Source: PB-295 001

Table 1.2: Summary Description of FGD Processes—Gypsum Category

FGD Process (year of first use)	Primary Reagent	Product for Disposal or Sale	Operating Principles	Present Status
Limestone/gypsum (1973)	$CaCO_3$	Gypsum	Flue gases scrubbed with limestone slurry. Resulting calcium-sulfur compounds are air-oxidized to gypsum, separated and dewatered.	Commercial operation (coal and oil), principally in Japan.
Lime/gypsum (1964)	CaO	Gypsum	Flue gases scrubbed with slurry of hydrated lime. Resulting calcium-sulfur compounds are air-oxidized to gypsum, separated and dewatered.	Commercial operation (coal and oil), principally in Japan; also recently in West Germany.
Double alkali/gypsum (1973)	NaOH or Na_2CO_3 and CaO or $CaCO_3$	Gypsum	Flue gases scrubbed with alkaline sodium compounds in solution. Used solution regenerated with lime or limestone in separate reactor and the insoluble calcium-sulfur compounds are air-oxidized to gypsum, separated and dewatered.	Commercial operation on oil-fired boilers in Japan.
Dilute sulfuric acid/gypsum (Chiyoda) (1972)	H_2SO_4 and $CaCO_3$	Gypsum	Flue gases are scrubbed with dilute sulfuric acid containing ferric ion catalyst. Air oxidation produces more sulfuric acid, which is bled off and neutralized with limestone. The gypsum is separated and dewatered.	Commercial operation on oil-fired boilers in Japan.

Source: PB-295 001

Table 1.3: Summary Description of FGD Processes—Regenerative Category

FGD Process (year of first use)	Primary Reagent	Primary Product	Operating Principles	Present Status
Sodium sulfite (Wellman-Lord) (1971)	Na_2CO_3 NaOH	SO_2	Flue gases are scrubbed with sodium sulfite solution and the resulting sodium bisulfite is thermally regenerated to sodium sulfite and SO_2. The SO_2 is then further treated to yield marketable products such as sulfuric acid or sulfur.	Commercial operation on oil-fired boilers in Japan. Recent commercial application on coal-fired boilers in U.S. Also on chemical/oil refinery plants.
Magnesium oxide (1971)	MgO	SO_2	Flue gases scrubbed with slurry of hydrated magnesium oxide. Insoluble magnesium sulfite is separated and calcined for recycle of MgO and generation of SO_2 for further treatment.	Commercial operation on oil-fired boilers in Japan. Demonstration plant on coal-fired boiler in U.S.
Carbon adsorption (1974)	Carbon	SO_2	SO_2 from flue gases is adsorbed on activated carbon, which is then regenerated thermally, releasing SO_2 for recovery.	Demonstration
Copper oxide (Shell) (1973)	CuO	SO_2	SO_2 from flue gases is adsorbed on copper oxide coated alumina. This is then treated chemically to regenerate the adsorbent and produce SO_2 for recovery.	Demonstration

Source: PB-295 001

Table 1.4: Summarized Data on FGD Systems Installed on Boiler Plants

FGD Process	Number of Systems Surveyed	MWe Range[a]	Predominant Fuel Type	Predominant Fuel Sulfur, %	Reported SO_2 Removal Efficiency Range[b], %	Reported Plant Availability Range[b], %	Process Capacity Range, Nm^3/s[c]
Limestone/sludge	3	115-820	Coal	0.5-5.4	57-92	88-94	82-108
Lime/sludge	6	65-835	Coal	0.8-4.7	70-92	83-100	28-139
Double alkali/sludge	2	20-32	Coal	0.9-5.0	90-99	80	25-35
Seawater scrubbing	1	(50)[d]	Oil	3.0	90-97[e]	96-99	17
Limestone/gypsum	6	250-850	Coal	0.6-2.2	90-96	97-100	103-250
Lime/gypsum	2[f]	37-750	[g]	2.9	80-92	96-98	35-333
Double alkali/gypsum	4	26-450	Oil	2.0-2.9	90-95	95-98	23-355
Dilute sulfuric acid/gypsum	2[f]	20-350	[g]	3.6[h]	96	100	270
Sodium sulfite	2	115-220	[g]	2.5-3.5	90-92	97-99	99-172
Magnesium oxide	4	95-155	[g]	2.0-3.8	90-95	i	102
Carbon adsorption	2	20-49	Coal	0.9-5.0	70-80	NA	10-42
Copper oxide	1	36	Oil	2.5	80-90	NA	30

[a] Range of MWe units served by FGD system

[b] Whole numbers

[c] Range of individual scrubber (or equivalent absorber) capacity

[d] Equivalent MWe

[e] Efficiency dependent on bypass (reheat) flow

[f] Includes one coal-fired demonstration unit

[g] Equal number of coal and oil firing systems were surveyed

[h] Calculated value

[i] Insufficient data base

NA: Not available

Source: PB-295 001

By late 1977, 92 systems were operating, of which 91 were in Japan and 1 in West Germany. They have been used successfully on both coal- and oil-fired plants and are reported to have high SO_2 removal efficiencies and high availabilities.

The gypsum processes have been applied particularly in Japan. Production of FGD-gypsum and phosphate fertilizer-gypsum has led to an annual oversupply of about 2 x 10^6 t/y. This gypsum oversupply and the reportedly lower cost of low sulfur fuel (desulfurized fuel oil and low sulfur crudes) are mentioned as reasons why no new gypsum producing FGD plants will be built on oil-fired boilers.

As with the throwaway process, the double alkali and the dilute sulfuric acid processes were developed to overcome some of the operating problems originally present with the lime or limestone scrubbing processes.

Regenerative Processes: Interest in the two surveyed regenerative processes that use wet scrubbing of the flue gases, i.e., the sodium sulfite and magnesium oxide processes, is largely due to the regeneration of the absorbent and production of a saleable sulfur product. No additional land area for solid waste disposal is required (beyond that for ash disposal), and raw material consumption is greatly reduced.

These processes have been successfully operated on oil-fired plants and are now being applied to coal-fired boilers. High SO_2 removal efficiencies are reported. With one of the processes (magnesium oxide), there is insufficient data to report on availability, but high availability is reported for the two Wellman-Lord systems surveyed for short periods.

The dry adsorption regenerative processes (carbon adsorption and copper oxide) are being investigated because of their potential for recovery of a sulfur by-product without the disadvantages of wet scrubbing systems, and with less energy requirement for flue gas reheat. These processes have not yet been applied on a commercial scale.

All of the regenerative processes require additional processing plants for converting the concentrated SO_2 into saleable form, e.g., by purification and liquefaction, by conversion into sulfuric acid, or by chemical reduction to elemental sulfur. The processes may be more suitable for operation near industrial areas where the sulfur products may be marketed. The wet scrubbing processes also offer the possibility of utilizing a single, centralized regeneration facility to service a number of FGD systems.

APPLICATION AND PERFORMANCE OF FGD TECHNOLOGY IN THE UNITED STATES

> The information in this section is based on "Status of Flue Gas Desulfurization in the United States," prepared by B.A. Laseke and T.W. Devitt of PEDCo Environmental, Inc., in *Proceedings: Symposium on Flue Gas Desulfurization, Las Vegas, Nevada, March 1979,* Vol. I, EPA-60017-79-167a, compiled by F.A. Ayer of Research Triangle Institute for the U.S. Environmental Protection Agency. These proceedings will be hereinafter referred to as Symposium—Vol 1.

Significant development of FGD technology in the United States dates from about the 1950s when bench-scale and limited pilot plant programs were initiated. Major pilot plant investigations first started in 1961, and between 1961 and 1978 more than 60 systems representing generating capacity of approximately 75 MW were investigated at the pilot plant level in the utility sector. Concurrent with later pilot plant investigations, prototype, demonstration, and full-scale systems were installed. The first commercial application of an FGD system on a utility boiler occurred in 1968. Since then, 62 systems representing a generating capacity of approximately 17,600 MW have been operated at the prototype, demonstration, and full-scale levels. As of the end of November 1978, 46 systems representing a generating capacity of 16,054 MW were in service, and another 98 systems representing 46,453 MW were under construction or planned.

Since the early investigations considerable progress has been made in the development of FGD technology, and FGD is now considered to be the most commercially developed means of continuous control of sulfur dioxide emissions from coal-fired boilers.

The evolution of FGD technology from the limited pilot plant level to the full-scale commercial level can be attributed to several general process design and application considerations. Although it is difficult to quantify the impact that many of these factors have had on the development of the technology, an attempt is made to identify some of the general contributing factors in the balance of this section.

Process Design Strategy: Several general tendencies are evident in recent FGD design strategies. Generally, system designs incorporate an increased degree of flexibility and reliability. Specifically, trends are toward the paring of modules and ancillary components and the designing of less interdependent systems (i.e., systems in which major unit operations are not strongly affected by upstream component performance).

System Applications: Many recent FGD facilities are installed on large base-loaded units designed to fire coal from a specific source. This generally results in a flue gas with more constant and stable characteristics, which can improve system reliability because the system does not have to respond to as dramatic a variation in flue gas flow rate and composition. In many of the original FGD applications, the systems were required to operate on widely varying loads (cycling and peak) and coal types (low sulfur western, high sulfur eastern, and blends); such situations often demanded response to conditions beyond their process control capability. As a result, variations in the reagent feed rate, loss of chemical control, and the incidence of chemical and mechanical problems caused numerous forced outages and lower dependabilities.

System Supplier and Architectural-Engineering Experience: Later FGD system designs have benefited from experience gained in the operation of first-generation systems. Building on this experience, system suppliers and designers are providing better process design configurations and materials of construction. That many suppliers now offer broader quarantees covering sulfur dioxide removal, particulate loading, mist loading, waste stream quality/quantity, power consumption, water consumption, reagent consumption, reheat energy consumption, and availability is indicative of this trend.

Utility Experience: The utilities have also been gaining valuable operating and design experience. Many utilities have conducted or participated in FGD pilot plant programs and are thus better prepared to operate demonstration and full-scale systems. Operation of their first demonstration or full-scale system has also led to improved design and operation of subsequent systems.

Regulatory Agency Attitudes: As FGD technology has evolved from a research, development, and demonstration effort to a means of continuous compliance with applicable regulations, local, state, and Federal regulatory agencies have changed their attitudes toward enforcement, compelling utility companies to improve the reliability of FGD systems.

Process Chemistry: Although scale and corrosion are still encountered and are sometimes still severe, general knowledge concerning scale formation and the occurrence of corrosion has greatly improved. As a result, systems are being designed and operated so that problems experienced by the earlier units will not be encountered.

CONVENTIONAL PROCESSES

> The information in this section is based on two appendices to the NATO-CCMS study, *Flue Gas Desulfurization Pilot Study–Phase I–Survey of Major Installations: Appendix 95-A–Limestone/Sludge Flue Gas Desulfurization Process,* January 1979, NTIS PB-295 002, prepared by F. Princiotta of the U.S. Environmental Protection Agency and R.W. Gerstle and E. Schindler of PEDCo Environmental and *Appendix 95-B–Lime/Sludge Flue Gas Desulfurization Process,* January 1979, NTIS PB-295 003, prepared by N. Haug of Umweltbundesamt and G. Oelert and G. Weisser of Battelle-Institut e.V.

Limestone/Sludge Process

General Process Description: The principles of all limestone scrubbing systems are essentially the same. When the limestone-water slurry comes in contact with flue gas containing SO_2, the SO_2 is absorbed into the slurry and reacts with the limestone to form an insoluble sludge. The by-products include gypsum ($CaSO_4 \cdot 2H_2O$) and calcium sulfite hemihydrate ($CaSO_3 \cdot \frac{1}{2}H_2O$). These sludge by-products are generally disposed of in a pond. Figure 1.1 is an example of a flow diagram of a 500-MW, coal-fired boiler with a limestone/sludge FGD system.

Process Chemistry: The overall reactions that take place in the absorber are:

(1) $$SO_2 + CaCO_3 \rightarrow CaSO_3 + CO_2$$

(2) $$SO_3 + CaCO_3 \rightarrow CaSO_4 + CO_2$$

Many intermediate steps also take place, however. The calcium ion is formed during slurry preparation.

(3) $$CaCO_3(s) \rightarrow CaCO_3(aq)$$

(4) $$CaCO_3 + H_2O \rightarrow Ca^{++} + HCO_3^- + OH^-$$

The SO_3 anion forms at the flue gas-slurry interface in the absorber.

(5) $SO_2(g) \rightarrow SO_2(aq)$

(6) $SO_2(aq) + H_2O \rightarrow H_2SO_3 \rightarrow HSO_3^- + H^+$

(7) $HSO_3^- \rightarrow H^+ + SO_3^=$

The sulfite ion (equation 7) then combines with the calcium ion (equation 4) to form the precipitate calcium sulfite hemihydrate (equation 8).

(8) $Ca^{++} + SO_3^= + \frac{1}{2}H_2O \rightarrow CaSO_3 \cdot \frac{1}{2}H_2O(s)\downarrow$

Gypsum, an additional precipitate, is formed as follows:

(9) $SO_3^= + \frac{1}{2}O_2 \rightarrow SO_4^=$

(10) $Ca^{++} + SO_4^= + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O(s)\downarrow$

As reactions (8) and (10) proceed, the calcium cation is depleted from solution and additional $CaCO_3$ dissolves to react with the sulfite ion. In a limestone-sludge system, by-products occur from both reactions (8) and (10).

According to the molecular weights of limestone and SO_2, the theoretical requirement is 1 mol of limestone per mol of SO_2 removed. If a 20% excess stoichiometric amount and 95% purity of limestone are assumed, actual limestone required is 1.97 kg/kg of SO_2.

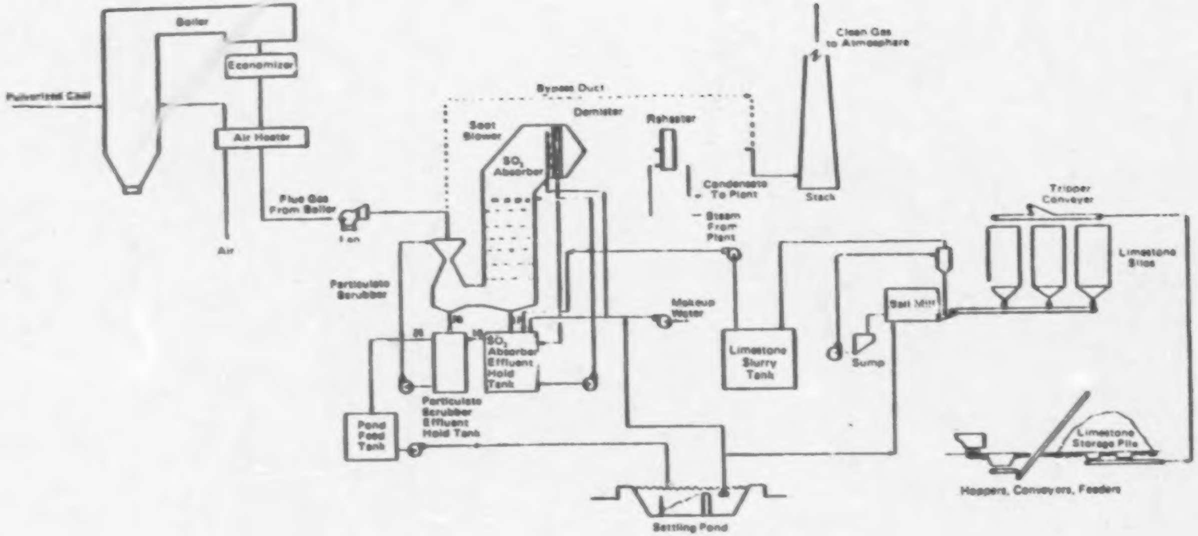
Dry sludge generated in the limestone process consists of calcium sulfite hemihydrate, carbonates, fly ash, and gypsum. Unused limestone and limestone impurities also combine with the sludge. The exact proportions of calcium sulfite hemihydrate and gypsum depend on system design; but if equal proportions are assumed, the sludge generated is 2.76 kg/kg of SO_2. When a venturi scrubber removes particulate matter, the particulates thus removed are also combined with the sludge. The final sludge to be disposed of contains at least 20% free water.

Description of Equipment Components: *Primary Particulate Removal* – The venturi scrubber is the first unit in most limestone FGD systems. This unit scrubs particulate from the gas; however, some SO_2 removal also occurs. The venturi scrubber has an advantage over an electrostatic precipitator (ESP) for particulate removal because the venturi cools and humidifies the flue gases before they enter the absorber section. A flue-gas cooler and humidifier must be used in connection with an ESP to cool the flue gases, generally to 50°C, prior to absorption. Moreover, if particulate is removed before the absorber, corrosion problems are reduced. Booster fans are sometimes installed in series with the venturi to provide the power necessary to force the gas through the scrubber system.

SO_2 Absorber – The absorber is the primary SO_2 removal unit in the system. Each of the many available designs employs a different method to contact the flue gas with the slurry. The most common unit designs include fixed packing, mobile-bed packing (hollow or solid spheres), and horizontal or vertical spray towers. Although each unit performs differently identical parameters have the same general effect on performance. Because of its simplicity, however, the spray tower is gaining popularity.

The scrubber must be constructed of materials that resist corrosion, erosion, and scaling. Scrubber bodies are fabricated of stainless steel or mild steel lined with an acid-resistant coating such as fiber glass-reinforced polyester (FRP), rubber or glass flake. Scrubber internals are made of a variety of materials such as stainless steel, which has a tendency to pit; high nickel alloys, which are expensive; or FRP, which is fragile. No one material seems to stand above the others.

Figure 1.1: Coal-Fired Boiler and Limestone Sludge FGD System


Pulverized Coal
Boiler
Economizer
Air Heater
Air
Flue Gas From Boiler
Fan
Particulate Scrubber
Bypass Duct
Soot Blower
SO_2 Absorber
Demister
Reheater
Condensate To Plant
Steam From Plant
Clean Gas to Atmosphere
Stack
Makeup Water
SO_2 Absorber Effluent Hold Tank
Particulate Scrubber Effluent Hold Tank
Pond Feed Tank
Settling Pond
Limestone Slurry Tank
Sump
Ball Mill
Tripper Conveyer
Limestone Silos
Limestone Storage Pile
Hoppers, Conveyors, Feeders

Based on Tennessee Valley Authority report, "Detailed Cost Estimate for Advanced Effluent Desulfurization Process."

Source: PB-295 002

The size and number of modules in a scrubber system are directly related to boiler size, turndown (reduction in boiler output) requirements, system availability, and gas-liquid distribution. Boiler system loads fluctuate, and the scrubber system must change to maintain optimum scrubber performance. One method of adjusting to turndown is to shut down scrubber modules as the load decreases. The more modules in the system, the smoother the transition. Scrubber modules not being used can be scheduled for cleaning and maintenance during periods of low system load, thereby reducing overall scrubber downtime. The use of multiple modules also has the advantage of permitting the modules to be smaller. Smaller cross-sectional areas in the scrubber module promote uniform gas-liquid distribution and improve efficiency. Scrubber module sizes range from about 25 to 200 MW.

Demister — A demister is necessary to remove entrained droplets from the scrubber outlet gas to reduce downstream equipment corrosion and scaling and to reduce reheat requirements. Most of the droplets are large enough to be removed with a simple change in flue gas direction; this is provided by baffles. Two banks of demisters are usually sufficient, but more can be added for additional demisting capability. Demisters are also subject to scaling and corrosion. Soot blowers or spray washes are installed to reduce this tendency, and materials of construction must be carefully selected.

Chevron and cyclonic-vane-type demisters are both effective. Although either can be operated in a horizontal or vertical position, the horizontal position predominates in the United States. There is much discussion pertaining to the advantages of each position, and a final selection depends on site-specific considerations.

Reheater — Reheating of stack gas is generally necessary to increase its bouyancy and reduce downstream corrosion. Thus, reheat not only helps meet ambient air standards; it also protects downstream equipment and prevents formation of acid mist. Reheating can be accomplished by installing a gas or low sulfur oil burner that exhausts directly into the stack, installing steam coils to heat the flue gas, injecting preheated air into the stack, or by-passing some hot flue gas around the FGD system directly into the stack (increasing emissions of SO_2). In-line heat exchangers are the most popular because of their low initial capital cost, but they tend to corrode and scale. Soot blowers, better demisters, and better materials of construction reduce these problems.

Slurry Makeup — Limestone can be received in a crushed and milled state or can be crushed and milled on site. In the latter case, the limestone is ground (wet or dry) in a ball mill to a size not larger than 200-mesh and often finer than 325-mesh. Finer grinding reduces the amount of limestone that remains unreacted and would otherwise be disposed of in the sludge. Water is added until the solids content reaches 15 to 25%. The slurry is sent first to a feed tank, then to the absorber holding tank, where it is mixed with absorber effluent. The slurry from the absorber holding tank is pumped to the absorber, where it reacts with SO_2 in the flue gas and is then returned to the holding tank. Slurry from the absorber holding tank is pumped to the venturi holding tank and from there to the venturi to scrub out fly ash. The slurry containing the fly ash returns to the venturi holding tank, from which it is pumped to the sludge disposal area for final treatment. Other designs have only one holding tank for the venturi and absorber.

Sludge Disposal — Sludge disposal can require 200,000 m^2 at a small plant and as much as 4,000,000 m^2 at a large plant. Disposal practices are very site specific. A power plant in an arid location might pump the sludge into an unlined pond, allowing the water to evaporate or seep into the ground. In an area where surface runoff or leaching could be a problem, the sludge sometimes is dewatered before being pumped into a lined pond. The water is returned to the system or purged after treatment to reduce chloride ions in the slurry. Plants located in areas where land is relatively expensive dewater the sludge to at most 20% liquid level by using clarifiers, thickeners, and vacuum filters and then fix the sludge by adding a proprietary compound. Fixation is necessary to allow the sludge pond to be used as a construction site at a later date. A totally closed system where no water is purged into a nearby stream or river has been achieved at the Sherburne County Station, Becker, Minnesota. Closed-loop systems are more subject to chloride ion concentration buildup in the slurry, especially when coal containing chloride is burned.

Advantages and Disadvantages: Limestone/sludge systems are the least expensive of all flue gas desulfurization systems currently used in the United States. Limestone is cheap and generally locally available, as is the land necessary for sludge disposal. In other countries, limestone appears to be available in sufficient quantities to make it inexpensive, but space for sludge disposal is generally not as available as in the United States.

The process is well developed chemically, but mechanical problems are still encountered in certain facilities. These problems include: fan vibration; pump and pipe erosion; scale buildup in the scrubber, demister and reheat sections; potential pollution in open-water systems; and corrosion and erosion. Essentially all these problems have been solved at particular installations. The similar limestone/gypsum systems in Japan offer the best examples of well-designed and well-operated limestone systems. Rapid response to changing boiler rates and flue gas conditions is essential in all limestone systems, and only highly trained chemical technicians or well-designed automatic controls can make the changes fast enough. The Japanese have succeeded in developing these automatic process controls on their limestone/gypsum systems.

The system operates well on large boilers fueled with either coal or oil. On small systems with low operating factors, labor and capital charges can be a limiting factor. Strict solid waste and water regulations either in force or imminent could necessitate more careful consideration of sludge disposal approaches. It may be necessary to incorporate an oxidation step and/or employ sludge stabilization/fixation to produce acceptable materials for landfill disposal.

The advantages of the limestone/sludge systems are as follows:

(1) The basic process is fairly simple and has few process steps.

(2) Capital and operating costs are relatively low, and reserves of limestone are fairly abundant.

(3) SO_2 removal efficiencies can be as high as 95%.

(4) The two-stage treatment of flue gases permits removal of SO_2 and particulates.

(5) Many years of operating experience have led to a greater understanding of the basic principles of this process.

(6) The process has demonstrated successful performance, especially in the removal of SO_2 from coal-fired systems.

(7) Fly ash does not adversely affect the system.

The disadvantages/problems of the limestone/sludge systems are as follows:

(1) Large quantities of waste must be disposed of in an acceptable manner.

(2) If not designed carefully or operated attentively, limestone systems have a tendency toward chemical scaling, plugging, and erosion which can frequently halt its operation.

(3) The scrubber requires high liquid-to-gas (L/G) ratios necessitating large pumps with attendant electrical requirements.

(4) The sludge may have poor settling properties when it has high sulfite content. Forced oxidation or soluble Mg in the slurry have been shown to lower sulfite content.

Lime/Sludge Process

General Process Description: The principles of all lime scrubbing systems are essentially the same. When the lime-water slurry or solution comes in contact with flue gas containing SO_2, the SO_2 is absorbed into the slurry and reacts with the lime to form an insoluble sludge. The by-products include gypsum ($CaSO_4 \cdot 2H_2O$), calcium sulfite hemihydrate ($CaSO_3 \cdot \frac{1}{2}H_2O$), and calcium carbonate ($CaCO_3$). These sludge by-products are generally disposed of in a pond. The process is schematically shown in Figure 1.2.

Process Chemistry: *Chemical Reactions –*

(1) $$CaO + H_2O \rightarrow Ca(OH)_2$$

(2) $$Ca(OH)_2 + SO_2 \rightarrow CaSO_3 + H_2O$$

(3) $$CaSO_3 + SO_2 + H_2O \rightarrow Ca(HSO_3)_2$$

(4) $$Ca(HSO_3)_2 + Ca(OH)_2 \rightarrow 2CaSO_3 + 2H_2O$$

Sulfate formation (detrimental), scaling:

(5) $$2CaSO_3 + O_2 \rightarrow 2CaSO_4$$

Control Theory – Scrubbing liquor is a slurried mixture of calcium hydroxide and calcium sulfite in water. The pH of slurry entering the scrubber is 8 to 10; low pH can cause gypsum scaling; high pH can cause formation of carbonates. Reaction tank (if used) pH is 10 (addition of lime); addition of crystal seed causes homogeneous crystallization. Presence of MgO in the lime allows a subsaturated mode of operation and improves the SO_2-removal efficiency.

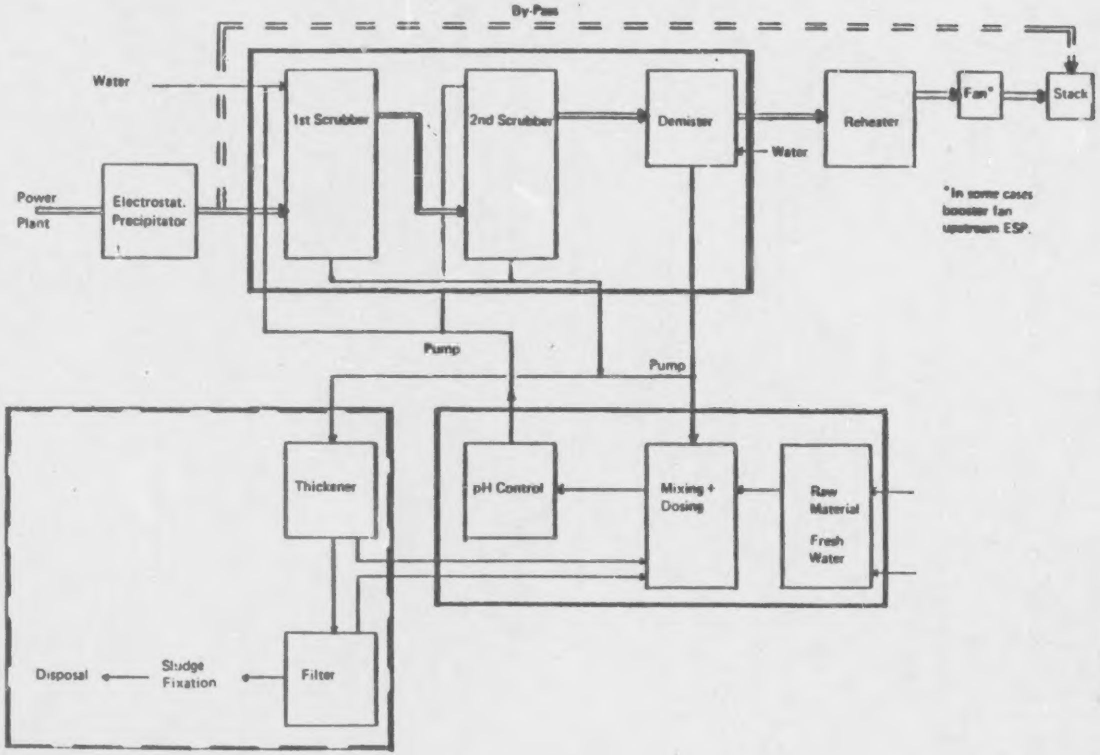
Reaction Kinetics – Reaction with SO_2 in the flue gas takes place in the liquid phase; dissolution of calcium sulfite is the rate controlling step for SO_2-absorption. In other cases the mass transfer through the interface between gas and liquid is the rate controlling step.

Stoichiometric Requirement – The equipment components are similar to those described for the limestone/sludge process.

Advantages and Disadvantages: Considering the application of lime/sludge flue gas desulfurization (FGD) systems, the space necessary for the sludge disposal is among others an important factor. Therefore, the lime/sludge (and the limestone/sludge) technology is economically advantageous in the U.S. due to the availability of inexpensive land, whereas in Europe and in Japan the availability of sufficient space is often prohibitive.

Generally inexpensive lime can be provided to the FGD plants and, as far as available, carbide sludge from chemical industry or alkaline fly ash can be utilized as scrubbing agent. The lime/sludge technology is well developed.

Figure 1.2: Flow Diagram—Lime/Sludge Process


By-Pass
Water
Power Plant
Electrostat. Precipitator
1st Scrubber
2nd Scrubber
Demister
Water
Reheater
Fan*
Stack
Pump
Pump
Thickener
pH Control
Mixing + Dosing
Raw Material
Fresh Water
Filter
Sludge Fixation
Disposal

*In some cases booster fan upstream ESP.

Source: PB-295 003

Through operation of large scale plants in different countries considerable experience was gathered and reliability of the systems was proven.

Current R&D efforts aim at the following chemical, mechanical and design areas:

- Precipitation of calcium sulfate (gypsum) may cause scaling, which is particularly unwanted in mist eliminators.
- Dissolved salts in the scrubbing agent and chloride built-up in the recycle water can cause corrosion, which is possibly aggravated by the erosive nature of the slurry.
- Pumps, fans and agitators allow mechanical improvements as to their use in this technology.
- Interrelated mechanical and chemical factors may influence the lifetime of expansion joints and piping.
- Finally the optimization of the design parameters like gas flow and slurry distribution, liquid-to-gas ratio, control instrumentation and accessibility for maintenance has to be mentioned.

The advantages of the lime/sludge system are similar to those listed for the limestone/sludge process.

The disadvantages and problems are also similar to those listed for the limestone/sludge process. In addition, although fly ash does not adversely affect the process in general it can adversely affect the process by intensification of mechanical wear and erosion in the washing cycle and by increased load of the thickener.

The limestone/gypsum and lime/gypsum processes described in Table 1.2 may also be considered conventional processes.

Dry FGD systems have been emphasized due to several problems associated with plugging and corrosion in wet processes. This Noyes report addresses the two major types of dry processes, and reviews the status of their development. A comparison with wet processes draws attention to the advantages of dry processes over the widely used wet processes.

DRY SYSTEMS

The information in this chapter is based on *Survey of Dry SO_2 Control Systems,* February 1980, EPA-600/7-80-030, by G.M. Blythe, J.C. Dickerman and M.E. Kelly of Radian Corporation for the U.S. Environmental Protection Agency.

INTRODUCTION

The purpose of this chapter is to summarize the status of dry flue gas desulfurization (FGD) processes in the United States, for both utility and industrial application. Throughout, dry FGD will be defined as any process which involves contacting a sulfur-containing flue gas with an alkaline material and which results in a dry waste product for disposal. This includes (1) systems which use spray dryers for a contactor, with subsequent baghouse or electrostatic precipitator (ESP) collection of waste products; (2) systems which involve dry injection of alkaline material into contact with flue gas, and subsequent baghouse or ESP collection; and (3) other varied dry systems which include concepts such as addition of alkaline material to a fuel prior to combustion or contacting flue gas with a fixed bed of alkaline material.

This definition of dry systems excludes several dry adsorption or "acceptance" processes, such as the Shell/UOP copper oxide process, or the Bergbau-Forschung adsorptive char process.

Also excluded is the regenerable Rockwell Aqueous Carbonate Process (ACP) which, although it does use a spray dryer for a flue gas contactor, does not fit the limitation of being a "throwaway" system. However, the open loop, spray dryer contactor portion of the Rockwell process has been adapted for a "throwaway" system and as such has been included here.

Technical Glossary: Definitions for several terms that are used frequently throughout this report to describe the operation of dry FGD systems are defined as follows:

Stoichiometry for dry scrubbing is defined as the mols of fresh sorbent introduced to the system divided by the mols theoretically required for complete reaction with all of the SO_2 entering the system whether or not it is all removed. This is opposed to wet scrubbing where stoichiometry is generally based on mols of SO_2 removed by the system.

Sorbent utilization is defined as the percent SO_2 removal by the system divided by the stoichiometry:

$$\frac{\left[\dfrac{\text{mols } SO_2 \text{ removed}}{\text{mols } SO_2 \text{ entering system}}\right] \times 100}{\left[\dfrac{\text{mols sorbent entering system}}{\text{mols sorbent required to react with } SO_2 \text{ entering}}\right]} = \text{percent utilization}$$

If one defines the sorbent for a calcium-based system as CaO and the sorbent for a sodium-based system as Na_2O, one mol of sorbent reacts with one mol of SO_x. Consequently, the above expression reduces to:

$$\% \text{ utilization} = \text{mols } SO_2 \text{ removed/mols sorbent entering system} \times 100$$

Since the mols of sorbent do not include alkalinity from other sources such as recycled fly ash, it is possible to see apparent utilizations of greater than 100 percent. That is, the alkalinity in recycled fly ash can react to remove SO_2 so that there are more mols of SO_2 removed than mols of fresh sorbent feed.

A spray dryer is defined as any apparatus in which flue gas is contacted with a slurry or solution such that the flue gas is adiabatically humidified and the slurry or solution is evaporated to apparent dryness. For FGD applications the material dried is often a calcium-based slurry or a sodium solution which reacts with flue gas sulfur during and following the drying process. The spray dryer can use rotary, two-fluid or nozzle atomization, and the vessel can be anything from the back-mix reactor typically used in spray dryer technology to a large horizontal duct.

Dry injection is defined as the process of introducing a dry sorbent into a flue gas stream. This can take the form of pneumatically injecting sorbent into a flue gas duct, precoating or continuously feeding sorbent into a flue gas duct, precoating or continuously feeding sorbent onto a fabric filter surface, or any similar form of mechanically introducing a dry alkaline sorbent into a flue gas stream.

Coal/limestone combustion is defined as the process of burning a mixture of coal and limestone whereby the SO_2 released from the coal reacts with the limestone to form solid calcium salts that are collected with the ash. Two specific combustion processes are discussed: one involves burning a coal/limestone pellet in a stoker-fired boiler, and the other involves burning a pulverized coal/limestone mixture in a low NO_x burner.

PROCESS ASSESSMENT

Spray Dryer-Based Systems

In these systems, flue gas at air preheater outlet temperatures (generally 250° to 400°F) is contacted with a solution or slurry of alkaline material in a vessel of relatively long residence time (5 to 10 seconds). The flue gas is adiabatically humidified to within 50°F of its saturation temperature by the water evaporated from the solution or slurry. As the slurry or solution is evaporated, liquid-phase salts are precipitated and remaining solids are dried to generally less than one percent free moisture. These solids, along with fly ash, are entrained in the flue gas and carried out of the dryer to a particulate collection device. Reaction between the alkaline material and flue gas SO_2 proceeds both during and following the drying process. The mechanisms of the SO_2 removal reactions are not well understood, so it has not been determined whether SO_2 removal occurs predominantly in the liquid phase, by adsorption into the finely atomized droplets being dried, or by reaction between gas phase SO_2 and the slightly moist spray-dried solids.

Sodium carbonate solutions and lime slurries are common sorbents. A sodium carbonate solution will generally achieve a higher level of SO_2 removal than a lime slurry at similar conditions of inlet and outlet flue gas temperatures, SO_2 level, sorbent stoichiometry, etc. Lime, however, has become the sorbent of choice in many circumstances because of the cost advantage it enjoys over sodium carbonate and because the reaction products are not as water-soluble. Through the use of performance-enhancing process modifications, such as sorbent recycle and hot or warm gas bypass, lime sorbent has been demonstrated at the pilot scale to achieve high levels of removal (85 percent and greater) at sorbent utilization near 100 percent.

Using a spray dryer for a flue gas contactor involves adiabatically humidifying the flue gas to within some approach to saturation. With set conditions for inlet flue gas temperature and humidity and for a specified approach to saturation temperature, the amount of water which can be evaporated into this flue gas is set by heat balance considerations. Liquid to gas ratios are generally in the range of 0.2 to 0.3 gal/Mcf. The sorbent stoichiometry is varied by raising or lowering the concentration of a solution or weight percent solids of a slurry containing this set amount of water. While holding other parameters such as temperature constant, the obvious way to increase SO_2 removal is to increase sorbent stoichiometry.

However, as sorbent stoichiometry is increased to raise the level of SO_2 removal, two limiting factors are approached: (1) sorbent utilization decreases, raising sorbent and disposal costs on the basis of SO_2 removed; and (2) an upper limit is reached on the solubility of the sorbent in the solution, or on the weight percent of sorbent solids in a slurry.

There are at least two methods of circumventing these limitations. One method is to initiate sorbent recycle, either from solids dropped out in the spray dryer or from the particulate collection device catch. This has the advantage of increasing the sorbent utilization, plus it can increase the opportunity for utilization of any alkalinity in the fly ash.

The second method of avoiding the above limitations on SO_2 removal is to operate the spray dryer at a lower outlet temperature, that is, a closer approach to saturation. Operating the spray dryer outlet at a closer approach to saturation has the effect of both increasing the residence time of the liquid droplets and increasing the residual moisture level in the dried solids. As the approach to saturation is narrowed, SO_2 removal rates and sorbent utilization generally increase dramatically. Since the mechanisms for SO_2 removal do not appear to be well understood, it is not obvious whether it is the increase in liquid phase (droplet) residence time, or the increase in residual moisture in the solids, or both which accounts for the increased removal.

Unfortunately, the approach to saturation at the spray dryer outlet is set by either the requirement for a margin of safety to avoid condensation in downstream equipment or restrictions on stack temperatures. The spray dryer outlet can be operated at temperatures lower than these restrictions would otherwise allow if some warm or hot gas is bypassed around the spray dryer and used to reheat the dryer outlet. Warm gas (downstream of the boiler air heater) can be used at no energy penalty, but the amount of untreated gas involved in reheating begins to limit overall SO_2 removal efficiencies. Significantly less hot gas (upstream of the air heater) is required to heat, but an energy penalty associated with the decrease in heat load to the air heater comes with bypassing the plant air heater. Figure 7.1, a general flow diagram of a spray dryer-based system, illustrates these two "reheat" options.

The spray dryer design can be affected by the choice of particulate collection device. Bag collectors have an inherent advantage in that unreacted alkalinity in the collected waste on the bag surface can react with remaining SO_2 in the flue gas. Some process developers have reported SO_2 removal on bag surfaces on the order of 10 percent (1). A disadvantage of using a bag collector is that since the fabric is somewhat sensitive to wetting, a margin above saturation temperature (on the order of 25° to 35°F) must be maintained for bag protection. ESP collectors have not been demonstrated to achieve significant SO_2 removal. However, some vendors claim that the ESP is less sensitive to condensation and hence can be operated closer to saturation (less than a 25°F approach) with the associated increase in spray dryer performance.

The choice between sorbent types, use of recycle, use of warm or hot gas bypass, and types of particulate collection device tends to be rather site specific. Vendor and customer preferences, system performance requirements, and site-specific economic factors tend to dictate the system design for each individual application.

Dry Injection Process

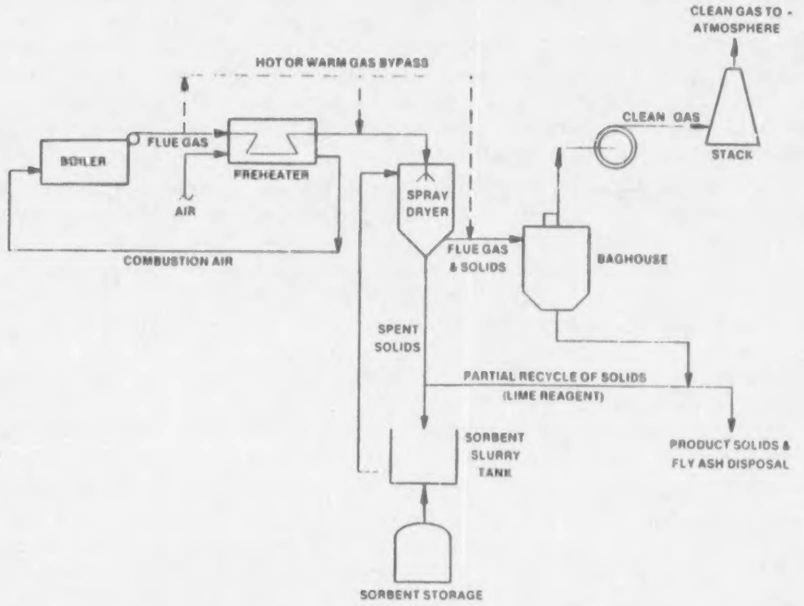
Dry injection schemes generally involve pneumatically introducing a dry, powdery alkaline material into a flue gas stream with subsequent particulate collection. A generalized flow diagram of this process is shown in Figure 7.2. The injection point has been varied from the boiler furnace area all the way to the flue gas entrance to an ESP or bag collector. Most dry injection schemes use a sodium-based sorbent. Lime has been tested but has not been demonstrated with much success. Many dry injection programs have used nahcolite as a sorbent. Nahcolite is a naturally occurring mineral, associated with western oil shale reserves, and is about 80 percent sodium bicarbonate. Sodium bicarbonate appears to be more reactive than sodium carbonate, because it loses two mols of CO_2 and one mol of water in reaction, while sodium carbonate loses only one mol of CO_2 in reaction with SO_2. The following overall reactions illustrate this point:

$$2NaHCO_3 + SO_2 \rightarrow Na_2SO_3 + H_2O + 2CO_2$$

$$Na_2CO_3 + SO_2 \rightarrow Na_2SO_3 + CO_2$$

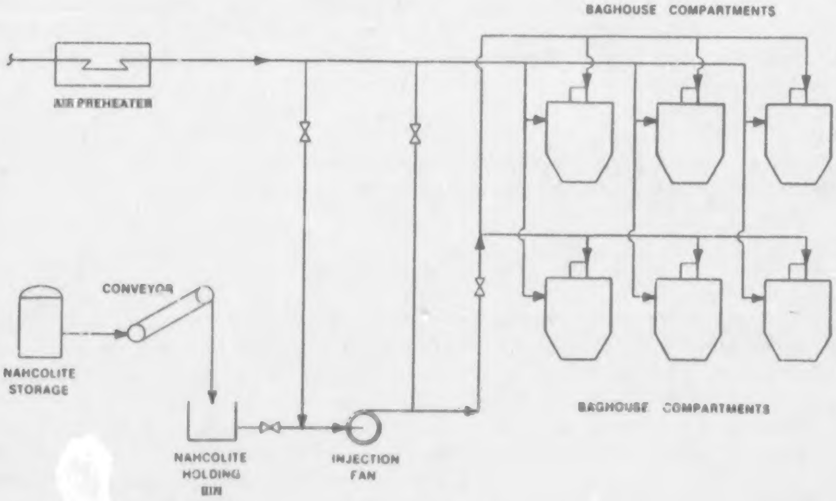
Since bicarbonate loses three mols for every mol of SO_2 removal, bicarbonate particles tend to have larger pore volumes and are apparently less susceptible to blinding on reaction than are sodium carbonate particles. Unfortunately, the availability of raw nahcolite in commercial quantities in the near future is questionable due to the substantial investment necessary before commercial-scale mining can begin. Since the potentially favorable economics of dry injection are based to some extent on the use of inexpensive sorbents, the use of commercially refined sodium bicarbonate is prohibitively expensive.

Figure 7.1: Typical Spray Dryer/Particulate Collection Flow Diagram


CLEAN GAS TO ATMOSPHERE
HOT OR WARM GAS BYPASS
CLEAN GAS
STACK
FLUE GAS
BOILER
PREHEATER
AIR
SPRAY DRYER
FLUE GAS & SOLIDS
BAGHOUSE
COMBUSTION AIR
SPENT SOLIDS
PARTIAL RECYCLE OF SOLIDS (LIME REAGENT)
SORBENT SLURRY TANK
PRODUCT SOLIDS & FLY ASH DISPOSAL
SORBENT STORAGE

Source: EPA-600/7-80-030

Figure 7.2: Nahcolite Dry Injection Flow Diagram


BAGHOUSE COMPARTMENTS
AIR PREHEATER
CONVEYOR
NAHCOLITE STORAGE
NAHCOLITE HOLDING BIN
INJECTION FAN
BAGHOUSE COMPARTMENTS

Source: EPA-600/7-80-030

Recent research has been aimed at studying the use of raw trona ore, which is currently mined in large quantities both in the Green River, Wyoming area and the Owens Lake, California area. The mineral trona contains one mol of sodium carbonate, one mol of sodium bicarbonate and two waters of hydration ($Na_2CO_3 \cdot NaHCO_3 \cdot 2H_2O$). Trona has the potential for providing a good compromise between reactivity, cost, and availability for use in dry injection schemes.

An unresolved problem with this technology is disposal of the sodium-based waste materials in an environmentally acceptable manner. Sodium waste materials are highly soluble and can result in contamination of aqueous streams. Disposal of sodium compounds is an area requiring further investigation.

Both baghouse and ESP collection devices have been tested with dry injection processes. However, the effect of the reaction between unspent sorbent on collecting bag surfaces and SO_2 remaining in the flue gas seems to overwhelmingly favor the bag collector (2). Since a major portion of the SO_2 removal reaction appears to take place on the bag surface, various methods of feeding have been tested:

(1) Continuous. After the bag is cleaned, sorbent entrained with the flue gas is added to the bag surface continuously from injection points located upstream of the baghouse.
(2) Batch. After the bag is cleaned, all sorbent is added to the bag as a precoat before flue gas flow is resumed.
(3) Semibatch. This feeding method is a compromise between types (1) and (2). After bag cleaning some sorbent is added initially as a precoat and the remainder is added continuously throughout the bag cycle from an upstream injection point.

Sorbent stoichiometry, sorbent particle size, point and temperature of injection, baghouse air-to-cloth ratio, and bag cleaning frequency are also varied in dry injection programs.

Combustion of Coal/Limestone Fuel Mixture

The current research on combustion of a coal/limestone fuel mixture has taken two forms:

(1) Combustion of a coal/limestone pellet in an industrial spreader-stoker boiler.
(2) Combustion of a pulverized coal/limestone mixture in a low-NO_x burner system.

Preliminary results of bench-scale test work on both processes have indicated that up to 80% of the available sulfur in the fuel can be retained by the limestone. The ratio of calcium to sulfur in the coal/limestone fuel mixture is important in determining how much sulfur is retained.

A spreader-stoker boiler (20 bhp) has been used in testing the combustion and sulfur retention characteristics of the coal/limestone pellet. A Ca:S mol ratio of 7:1 has been used so far, but further work with a 3:1 Ca:S pellet is planned. The emissions generated are dependent upon burner design, coal properties and combustion operating parameters. The inherent staged combustion of the stoker-fired boiler (accomplished by supplying the total combustion air as primary air through the grates and secondary air through over-fire jets above the bed) results in lower NO_x emissions relative to conventional pulverized coal-fired boilers.

The two-staged combustion concept was employed by Babcock & Wilcox (B&W) to design an advanced low-NO_x burner system. EPA has funded test work to develop a concept of firing a coal/limestone fuel mixture in B&W low-NO_x burners to reduce SO_2 emissions. Tests conducted on a 12 × 10^6 Btu/hr unit by the Energy and Environmental Research Corporation (EERC), with a Utah low-sulfur coal, have demonstrated 88 percent SO_2 removal with a 3:1 Ca:S mol ratio. This high SO_2 removal has been attributed to the lower flame temperature found in the low-NO_x burner which may help maintain limestone reactivity. EERC has reported that SO_2 removal increased substantially when the reagent was passed through the pulverizer with the coal.

Further research on a larger scale for both systems is needed to determine the effects of combustion of a coal/limestone fuel mixture on boiler operation and maintenance. Collection of the increased ash loading and investigation of the properties and disposal of the waste products must also be studied.

COMPARISON OF DRY AND WET SCRUBBING

Comparisons between dry and wet scrubbing systems can be drawn in five major areas: waste disposal, reagent requirements, operation and maintenance, energy requirements, and economics. These comparisons will focus on general aspects of dry FGD systems as compared to conventional lime/limestone wet scrubbing systems.

With regard to waste disposal, dry FGD systems have an inherent advantage over wet lime/limestone systems in that they produce a dry, solid waste product that can be handled by conventional fly ash handling systems, eliminating requirements for a sludge handling system. However, the waste solids from sodium-based dry FGD systems are quite water-soluble and can lead to leachability and waste stability problems. Waste solids from lime spray drying systems and coal/limestone fuel systems should have similar environmental impacts as waste from lime/limestone wet systems, for which waste disposal technology is better defined.

In general, dry FGD systems require a higher stoichiometric ratio of sorbent to entering SO_2 to achieve the desired removal efficiency than do conventional limestone wet scrubbing systems. In addition, the reagents employed in spray drying and dry injection systems (soda ash, lime, commercial and naturally occurring sodium carbonates and bicarbonates, such as nahcolite and trona) are significantly more expensive than limestone. Consequently, limestone wet scrubbing systems will have an advantage with regard to reagent utilization and sorbent-related operating costs.

It has been claimed that dry systems will have lower maintenance requirements than comparable wet systems. Dry systems require less equipment than wet systems as the thickeners, centrifuges, vacuum filters and mixers required to handle the wet sludge waste product from wet systems are eliminated. In addition, slurry pumping requirements are much lower for spray drying and are eliminated in dry injection and combustion of coal/limestone fuel systems. This is important because wet systems have reported high maintenance requirements associated with large slurry circulation equipment. Finally, the scaling potential in wet limestone systems requires extra effort to maintain proper scrubber operation and possibly makes dry systems somewhat more flexible as far as their ability to adjust process operations to respond to variations in inlet SO_2 concentrations and flue gas flow rates.

With regard to energy requirements, dry FGD systems appear to have a significant advantage over wet systems due to savings in reheat and pumping requirements. Many wet FGD systems reheat the flue gas before it enters any downstream equipment to prevent corrosion. This reheat requirement is eliminated in most dry system configurations and results in considerable energy savings. Spray dryer systems are usually designed with a 30° to 50°F approach to the adiabatic saturation temperature of the flue gas at the outlet of the spray dryer. Energy savings from reduced pumping requirements result from the fact that wet scrubbers may require liquid to gas (L/G) pumping rates of up to 100 gallons per 1,000 acfm of gas whereas spray dryers only require an L/G rate of 0.2 to 0.3.

One of the major driving forces for development of dry SO_2 removal systems is the opportunity for reduction in both capital and operating costs. Although costs are quite site specific, the three types of dry FGD technologies considered here offer several potential possibilities for cost savings. This is due to the reduction in equipment and operation and maintenance requirements relative to conventional wet lime/limestone systems, especially in utility applications. Basin Electric evaluated the costs of the two spray drying systems they have purchased (Antelope Valley and Laramie River Stations) to be about 20 to 30 percent less costly over the 35-year life of the plants than comparable wet systems. However, it should also be noted that these economics are based on pilot-scale data and should be better determined after the operation of commercial systems has begun. Reagent costs for sodium-based spray dryer systems will be considerably higher than for lime-based systems although vendors claim the capital costs, excluding waste disposal, will be lower for sodium-based systems.

The minimal equipment and operating requirements for dry injection systems make the process economically attractive as far as capital costs are concerned, but high sorbent requirements and uncertainties in sorbent availability and cost are slowing further development of the technology on a commercial scale. Capital costs for both the pellet and low-NO_x burner coal/limestone fuel mixture systems should also be low since they will consist mainly of the equipment needed to produce the mixtures. However, since these systems have the potential for impacting the design and/or operation of the boiler, more information on the overall operability of these systems is needed before overall operating costs can be estimated.

In summary, dry systems do offer potential advantages over wet systems, especially in the areas of energy savings and costs. However, crucial issues such as waste disposal and demonstration of commercial-scale systems, which may continue to limit the overall acceptability of this technology, remain to be answered.

Advanced FGD systems, developed in response to operational difficulties with initial wet processes, potentially offer greater dependability and reliability than previous systems. A general description of these advanced processes and their developer are given in this Noyes Data report.

SURVEY OF ADVANCED SYSTEMS

EMERGING PROCESSES

> The information in this section is based on "Status of Flue Gas Desulfurization in the United States," prepared by B.A. Laseke and T.W. Devitt of PEDCo Environmental, Inc. in Symposium—Vol 1.

Considerable progress has been made in the development of conventional and emerging or advanced FGD processes. Much of this information has been acquired from the design and operation of first-generation FGD systems and translated into more effective designs and improved operation of newer systems. This section summarizes these emerging processes and presents a brief overview of their current status.

For the purposes of this discussion, processes within the emerging or advanced category are defined as those that incorporate major design and operating changes and thereby differ significantly from conventional direct lime/limestone processes. Of the processes so categorized, several have been evaluated at pilot and prototype development levels, and a few have progressed to the installation and operation of demonstration or full-scale units. Table 3.1 provides a brief summary of the emerging processes and highlights their current level of development and the extent of operating experience.

As is evident from Table 3.1, most of the previous operating experience has been with wet-phase sodium- and magnesium-based processes that produce a recoverable, marketable by-product or a high-quality filter cake. As might be expected, this substantial experience has resulted in the commercial application of dual alkali, Wellman-Lord, sodium carbonate, and magnesium oxide scrubbing systems.

The most recent promising development in emerging processes involves dry collection systems. Integrated processes designed to remove two or more pollutants simultaneously from the flue gas of a coal-fired boiler have commanded considerable interest for economic operational reasons. The success of fabric filters in removing particulates from the flue gases of coal-fired boilers prompted investigations into the feasibility of using these filters to control both particulates and sulfur dioxide.

Table 3.1: Major Emerging FGD Processes

Process	Developer	Current Level of Development	Previous Operating Experience	Remarks
Aqueous carbonate	Rockwell International	100-MW system (planned)	Mohave pilot plant test program	Full-scale application not yet demonstrated.
Catalytic oxidation	Monsanto	100-MW system (terminated)	Wood River test program	No further process development.
Chiyoda Thoroughbred 101	Chiyoda International	20-MW system (terminated)	Scholz prototype test program	Development of the process has ceased in favor of a new design concept (Thoroughbred 121 which employs limestone reagent in a jet bubbler reactor).
Copper oxide absorption	Shell/Universal Oil Products	pilot plant	Big Bend test program	Process available for prototype or demonstration application. No systems planned at present.
Dual alkali	A.D. Little/Combustion Equipment Associates	277-MW system (construction)	Scholz prototype test program	Full-scale application not yet demonstrated.
	FMC	250-MW system (construction)	Industrial systems and utility pilot plants	Full-scale application not yet demonstrated.
	Buell/Envirotech	575-MW system (construction)	Gadsby pilot plant test program	Full-scale application not yet demonstrated
Magnesium oxide	Chemico	150-MW system (terminated) 95-MW system (terminated)	Mystic test program; Dickerson test program	Process demonstrated on full-scale oil- and coal-fired boilers. System now offered for commercial application.
	United Engineers	600-MW system (planned)	Eddystone test program	Eddystone 120-MW prototype test program still in progress. A full-scale 600-MW system planned for TVA's Johnsonville Station.
Sodium carbonate	A.D. Little/Combustion Equipment Associates	Three 115-MW systems (operational)	Reid Gardner Station	Three full-scale sodium carbonate (trona) FGD systems have been in service on coal-fired boilers at Reid Gardner Station (Nevada Power). System performance has been good.
	Universal Oil Products	509-MW system (planned)	Jim Bridger pilot test program	Full-scale sodium carbonate (30% sodium carbonate purge solution from soda ash plant) FGD system planned for Jim Bridger Station.

(continued)

Table 3.1: (continued)

Process	Developer	Current Level of Development	Previous Operating Experience	Remarks
Wellman Lord	Davy Powergas	115-MW system (operational) 375-MW system (operational) 340-MW system (operational)	Crane test program	115-MW NIPSCO/EPA test program in progress at Mitchell. Two full-scale systems have recently started operations at Public Service of New Mexico's San Juan Station.
Dry adsorption	Foster Wheeler-Bergbau Forschung	20-MW system (terminated)	Scholz prototype test program	No further development of system reported. Further evaluation of the sulfur reduction component (RESOX) in Germany.
Dry collection	Wheelabrator-Frye Rockwell International	410-MW system (planned)	Leland Olds and Bowen Engineering pilot plant test program	Successful testing has resulted in planning of full-scale spray dryer and fabric filter FGD system at the Coyote Station.
	Joy/Niro	455-MW system (planned)	Leland Olds pilot plant test program	Successful pilot plant testing has resulted in planning of full-scale system at Antelope Valley. This system will use lime slurry as the reagent in 2-stage atomizer/fabric filter design configuration.
	Babcock & Wilcox	550-MW system (planned)	Basin electric pilot plant test program	Successful pilot plant testing has resulted in planning full-scale system at Laramie River. This system will use lime slurry as reagent in a configuration which employs an ESP as the second stage collector.
	Carborundum/ Delaval	pilot plant	Leland Olds pilot plant test program	Process similar in design to Wheelabrator-Frye/R.I. and Joy/Niro processes. No full-scale applications have been announced.

Source: Symposium-Vol 1

PROCESS DESCRIPTIONS

The information in this section is based on the following reports:

Economic and Design Factors for Flue Gas Desulfurization Technology, April 1980, EPRI CS-1428, prepared by Bechtel National, Inc. for Electric Power Research Institute;

Comparative Economics of Advanced Regenerable Flue Gas Desulfurization Processes, March 1980, EPRI CS-1381, prepared by M.R. Beychok for Electric Power Research Institute;

"Flue Gas Desulfurization and Fertilizer Manufacturing: Pircon-Peck Process" by R.B. Boyda of A.G. McKee & Company in *Proceedings: Symposium on Flue Gas Desulfurization, Las Vegas, Nevada, March 1979,* Vol II, EPA-60017-79-167b, compiled by F.A. Ayer of Research Triangle Institute for the U.S. Environmental Protection Agency (These proceedings will be hereinafter referred to as Symposium–Vol 2.);

"SO_2 and NO_x Removal Technology in Japan" by J. Ando of Chuo University in Symposium–Vol 2;

NATO-CCMS study, *Flue Gas Desulfurization Pilot Study–Phase I–Survey of Major Installations–Appendix 95-H, Flue Gas Desulfurization by Scrubbing with Dilute Sulfuric Acid,* January 1979, NTIS PB-295 009, prepared by F. Princiotta of U.S. Environmental Protection Agency and R.W. Gerstle and E. Schindler of PEDCo Environmental.

This section presents descriptions of advanced processes with only brief summaries included of those processes more fully described in subsequent chapters.

Process Summaries

Double Alkali Process: As applied to U.S. FGD applications, a double alkali process utilizes a sodium sulfite solution to absorb SO_2 from flue gas. The spent absorbent is reacted with lime to precipitate calcium sulfite and regenerate the active sodium absorbent. The precipitated calcium salts are separated and dewatered for disposal.

There are two versions of the process, a concentrated mode and a dilute mode. The selection is dependent on the level of absorbent oxidation (sulfite conversion to sulfate) experienced during absorption. Oxidation is favored by high oxygen-to-sulfur-dioxide ratios and by any catalytic activity contributed by the traces of fly ash captured by the absorber.

The "concentrated" mode process is favored for high-sulfur-coal uses (low oxidation) and is the simpler of the two process versions. The "dilute" mode process is applicable to low-sulfur-coal cases (high oxidation) and is the more complex and inherently more costly process for the same scope.

Wellman-Lord Process: The Wellman-Lord process employs wet absorption of SO_2 from flue gas by reaction with sodium sulfite to form sodium bisulfite and some sodium sulfate. Desulfurized flue gas is reheated and released to the atmosphere. The primary reactions occurring in the absorber are:

$$(1) \qquad Na_2SO_3 + H_2O + SO_2 \rightarrow 2NaHSO_3$$

$$(2) \qquad Na_2SO_3 + \tfrac{1}{2}O_2 \rightarrow Na_2SO_4$$

The absorber effluent liquor is filtered to remove solids and divided into two streams, a portion going to the regeneration area and the remainder going to purge treatment to reject the unreactive sodium sulfate. Double-effect evaporator crystallizers are used to convert the dissolved $NaHSO_3$ to crystalline Na_2SO_3 and liberated SO_2 by the reverse of reaction (1). The regenerated Na_2SO_3 crystals are dissolved and returned to the absorbers. The regenerated SO_2 stream is converted to elemental sulfur by reduction with natural gas in an Allied Chemical SO_2 reduction plant.

Magnesia Slurry Process: The magnesia slurry process uses an aqueous slurry of magnesium oxide to absorb and remove SO_2 from flue gases. The resulting magnesium sulfite and magnesium sulfate slurry is dried and calcined to yield recycle magnesium oxide and an SO_2-rich gas suitable for conversion to sulfuric acid.

The magnesia slurry process has been demonstrated in operation at oil- and coal-burning power plants in the United States and in Japan. The demonstration plants have operated on flue gas flows equivalent to those from power plants with outputs of up to 150 MW. The process design, engineering and construction of proprietary versions of the magnesia slurry process can be obtained in the United States from the Chemico Air Pollution Control Company and from United Engineers and Constructors.

Citrate Process: The absorption/steam stripping process as marketed by Flakt, Inc., the United States subsidiary of AB Svenska Flaktfabriken (SD) of Sweden, utilizes a solution of sodium citrate to absorb SO_2 from flue gas. The scrubber effluent liquor is stripped with low-pressure steam to liberate SO_2. The dilute SO_2 stream is then reduced to elemental sulfur using anthracite coal in a Foster Wheeler RESOX plant. A portion of the stripped effluent liquor is treated to remove sulfates and impurities and then recombined with the balance of the liquor before being returned to the scrubber sprays. The overall chemical reactions occurring in this process may be written as follows:

Absorption

$$Na_3Ci + SO_2 + H_2O \rightarrow NaHSO_3 + Na_2HCi$$

$$Na_2HCi + SO_2 + H_2O \rightarrow NaHSO_3 + NaH_2Ci$$

$$NaH_2Ci + SO_2 + H_2O \rightarrow NaHSO_3 + H_3Ci$$

Oxidation

$$2NaHSO_3 + O_2 \rightarrow Na_2SO_4 + H_2SO_4$$

$$H_2SO_4 + Na_3Ci \rightarrow Na_2SO_4 + NaH_2Ci$$

Desorption

$$NaHSO_3 + Na_2HCi \xrightarrow{\text{Heat}} H_2O + SO_2 + Na_3Ci$$

Reduction

$$C + SO_2 \rightarrow CO_2 + S$$

Lime Slurry Spray Dryer/Fabric Filter Process: The lime slurry spray dryer/fabric filter process selected for evaluation in this report uses a slurry of hydrated lime [$Ca(OH)_2$], calcium sulfite, and fly ash to sorb SO_2 from flue gas by contact with atomized droplets of the alkali slurry. These droplets are typically generated by high-speed centrifugal atomizers.

The following reactions occur in the spray dryer reaction chamber:

(1) $Ca(OH)_2 + SO_2 \rightarrow CaSO_3 + H_2O$

(2) $CaSO_3 + \frac{1}{2}O_2 \rightarrow CaSO_4$

Reaction (1) is the primary reaction for SO_2 removal. The dry salt mixture produced by these reactions is usually about 70% anhydrous sulfite and 30% anhydrous sulfate.

The atomized droplets rapidly evaporate in the spray dryer to produce a cooled and partially humidified particulate-laden gas from which most of the SO_2 has been removed. The mixture of fly ash, reaction products, and unreacted absorbent is removed from the gas stream in fabric filters. As the flue gas passes through the particulate-laden fabric, some additional SO_2 is absorbed by the remaining unreacted alkali, reducing the total amount of unreacted alkali in the filter cake.

Clean flue gas discharges from the fabric filters (baghouses) to the ID fans which discharge directly to the chimney. Individual groups of filter bags are periodically isolated from the flue gas stream and back-blown with clean flue gas to dislodge the filter cake, which is discharged from the baghouse hoppers and conveyed to storage. The total waste product stream is divided, the bulk of the material being trucked to a landfill site, and a varying amount being recycled to the absorbent preparation section to utilize the unreacted alkali.

Dry Bicarbonate Process: The dry bicarbonate process, proposed by Atomics International (since renamed the Environmental & Energy Systems Division of Rockwell International), injects dry sodium bicarbonate into flue gases to react with and remove SO_2 from the gases. The dry reaction products (a mixture of sodium sulfate and unreacted sodium bicarbonate) are removed from the flue gases by passage through a baghouse. Subsequent processing of the reaction products yields regenerated sodium bicarbonate and hydrogen sulfide gas. The hydrogen sulfide gas is converted to elemental sulfur as an end product. The dry bicarbonate process was offered by Atomics International with United Engineers and Constructors participating as an architect-engineering subcontractor.

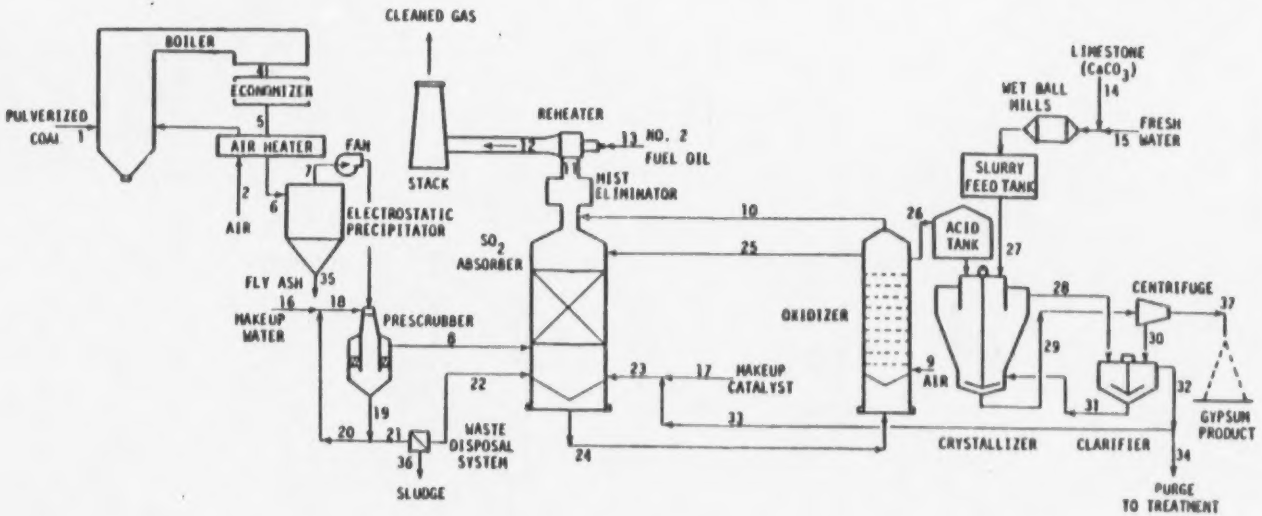
Carbon Adsorption Process: The dry adsorption process, developed by Bergbau Forschung GmbH removes SO_2 from flue gases by adsorption in a moving bed of activated char. Subsequent heating of the saturated char yields concentrated gaseous SO_2 and regenerates the char for reuse. The gaseous SO_2 is reduced and converted to elemental sulfur with the RESOX process, developed by the Foster Wheeler Energy Corporation, which uses anthracite coal to reduce the SO_2.

The overall combined system was jointly proposed by Bergbau Forschung GmbH, Foster Wheeler, STEAG AG and Deutsche Babcock AG with program support by the Federal Republic of Germany's environmental agency (Umweltbundesamt).

Chiyoda Thoroughbred 101 Process

General Process Description: The CT-101 process uses dilute sulfuric acid solution to absorb the SO_2 in the flue gas. The absorbed SO_2 gas is oxidized with air and forms additional sulfuric acid. The acid thus produced is reacted with limestone ($CaCO_3$) to produce gypsum ($CaSO_4 \cdot 2H_2O$) as a saleable by-product. Ferric ion was chosen as a catalyst for the oxidation step because it is harmless, available, and resistant to poisoning by contaminants. Figure 3.1 presents a simplified hypothetical flow diagram of a 500-MW, coal-fired boiler with a dilute sulfuric acid desulfurization system.

Figure 3.1: Flow Diagram for a Dilute Sulfuric Acid Desulfurization Process


CLEANED GAS
BOILER
ECONOMIZER
PULVERIZED COAL 1
4
5
AIR HEATER
2
AIR
FAN
7
6
ELECTROSTATIC PRECIPITATOR
STACK
12
REHEATER
13 NO. 2 FUEL OIL
11
MIST ELIMINATOR
SO_2 ABSORBER
FLY ASH 35
16
MAKEUP WATER
18
PRESCRUBBER
8
22
19
20
21
36
SLUDGE
WASTE DISPOSAL SYSTEM
24
10
25
23
17 MAKEUP CATALYST
33
OXIDIZER
9 AIR
26
ACID TANK
LIMESTONE ($CaCO_3$)
14
WET BALL MILLS
FRESH WATER 15
SLURRY FEED TANK
27
28
29
CRYSTALLIZER
CENTRIFUGE
37
30
31
32
CLARIFIER
34
PURGE TO TREATMENT
GYPSUM PRODUCT

Source: NTIS PB-295 009

Process Chemistry: The CT-101 process includes three major process areas: absorption, oxidation and crystallization. The overall reactions that take place in the absorber are:

$$SO_2\ (g) + H_2O \rightarrow H_2SO_3$$

At the flue gas–scrubbing solution interface in the absorber, the SO_3 anion forms in the following manner:

$$H_2SO_3 \rightarrow H^+ + HSO_3^- \quad \text{and} \quad HSO_3^- \rightarrow H^+ + SO_3^=$$

The catalytic oxidation reaction proceeds as follows:

$$SO_2\ (dis) + 2H_2O + Fe_2(SO_4)_3 \rightarrow 2FeSO_4 + 2H_2SO_4$$

$$SO_2\ (dis) + O_2\ (dis) + 2FeSO_4 \rightarrow Fe_2(SO_4)_3$$

$$2H^+ + SO_3^= + H_2O + Fe_2(SO_4)_3 \rightarrow 2FeSO_4 + 2H_2SO_4$$

In this stage, oxidation is aided by dissolved catalyst of ferric sulfate, $Fe_2(SO_4)_3$.

The sulfuric acid is sent to the crystallizer and neutralized with limestone to a concentration of 1% H_2SO_4, which generates a gypsum by-product in the following manner:

$$H_2SO_4 + H_2O + CaCO_3 \rightarrow CaSO_4{\cdot}2H_2O\downarrow + CO_2\uparrow$$

As a result of heat loss, the operating temperature is approximately 5°C lower than in the absorption and oxidation sections.

According to the stoichiometry, 1 mol of limestone is required to remove 1 mol of SO_2. Assuming 10% excess, actual limestone required is 1.72 kg/kg of SO_2 and the gypsum produced is 2.84 kg/kg of SO_2. The gypsum by-product contains about 10% free water.

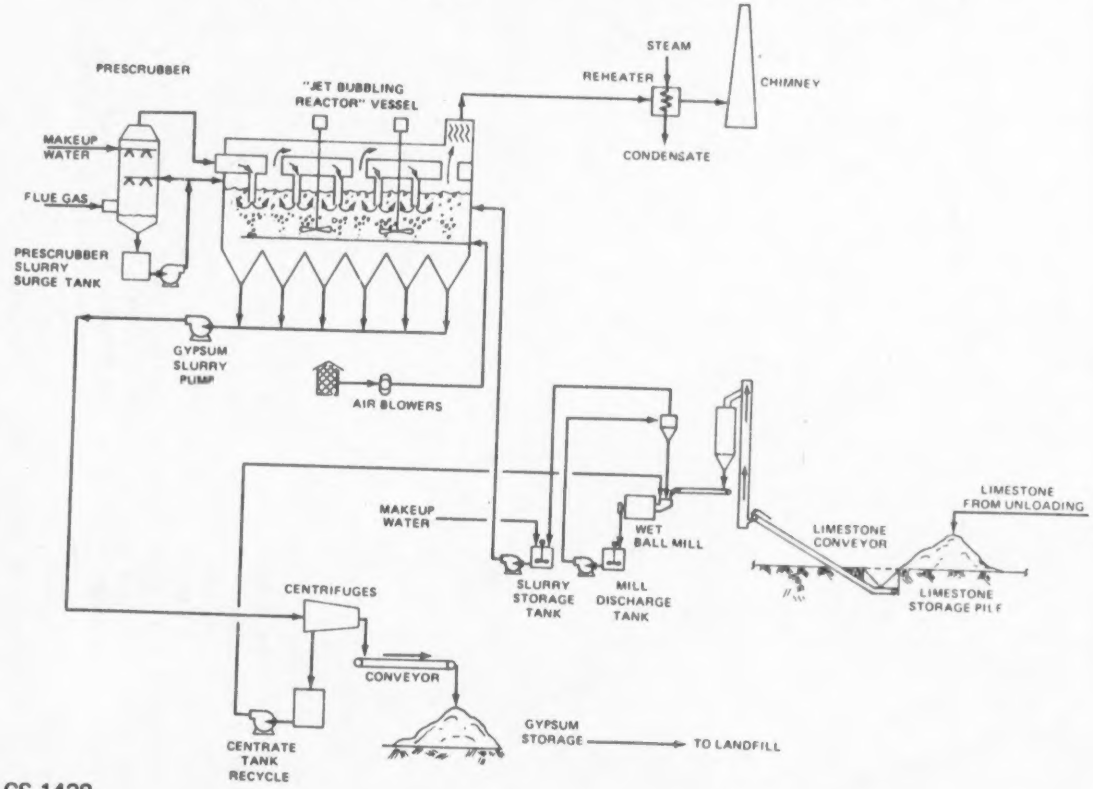
Chiyoda Thoroughbred 121 Process

Process Description: The Chiyoda Thoroughbred 121 (CT-121) process is a second-generation limestone slurry process in which SO_2 absorption, sulfite/bisulfite oxidation, and precipitation of calcium sulfate as gypsum are accomplished in a single reaction vessel. The overall chemical reactions are the same as those occurring in the limestone slurry process, except that virtually all of the sulfur is removed as an easily dewatered gypsum slurry rather than as mixed sulfite/sulfate sludge which is difficult to dewater. Figure 3.2 is a simplified flow diagram developed by Bechtel on the basis of previously published descriptions of the process.

A summary of the basic process design parameters for the CT-121 FGD system is presented in Table 3.2. The corresponding raw material and utility requirements are presented in Table 3.3. These data were provided by Chiyoda International Corporation, and were reproduced in EPRI CS-1428 with their concurrence.

Description of Major Process Subsystems: *Slurry Preparation* – Limestone is received by rail in uncovered bottom-dump rail cars. On-site storage consists of an uncovered reserve storage pile and a ball mill feed bin. Limestone is ground in wet ball mills and diluted with recycled process wastewater to produce slurry for makeup to the absorption section.

Figure 3.2: Chiyoda Thoroughbred 121 Process—Flow Diagram


PRESCRUBBER
"JET BUBBLING REACTOR" VESSEL
STEAM
REHEATER
CHIMNEY
CONDENSATE
MAKEUP WATER
FLUE GAS
PRESCRUBBER SLURRY SURGE TANK
GYPSUM SLURRY PUMP
AIR BLOWERS
MAKEUP WATER
WET BALL MILL
LIMESTONE CONVEYOR
LIMESTONE FROM UNLOADING
LIMESTONE STORAGE PILE
SLURRY STORAGE TANK
MILL DISCHARGE TANK
CENTRIFUGES
CONVEYOR
CENTRATE TANK RECYCLE
GYPSUM STORAGE
TO LANDFILL

Source: EPRI CS-1428

Table 3.2: Chiyoda Thoroughbred 121 Process System Parameters

Data for 1 x 500 MW Unit
(Maximum Load Conditions)

	Eastern Coal	Western Coal
Flue gas flow rate, m^3/sec (10^3 acfm)	856 (1,814)	942 (1,995)
Absorption trains per unit	4	4
Capacity per train, %	33⅓	33⅓
Gas flow per absorber, saturated, m^3/sec (10^3 acfm)	239 (507)	261 (552)
Presaturator type	.... spray tower.........	
Presaturator L/G, ℓ/m^3 (gal/10^3 acf)	0.936 (7)	0.936 (7)
Absorber type	......jet bubbling reactor.....	
Mist elimination	..........chevron..........	
System total pressure loss, pascals (inch wg)	4,982 (20)	2,740 (11)
Absorbent stoichiometric ratio*	1.01	1.01
Oxidation air stoichiometric ratio*, %	4	10
Flue gas reheat required, °C (°F)	31.1 (56)	26.7 (48)

*Mols alkali per mol SO_2 removed.

Source: EPRI CS-1428

Table 3.3: Chiyoda Thoroughbred 121 Process Raw Materials and Utility Requirements

Data for 1 x 500 MW Unit
(Maximum Load, Average Sulfur)

	Eastern Coal	Western Coal
Raw materials		
Limestone, 10^3 kg/hr (tons/hr)	25.6 (28.2)	3.9 (4.3)
Utilities		
Electricity		
Raw material handling and preparation, kW	730	110
Pretreatment and absorption, kW	2,510	510
Waste dewatering and storage, kW	300	55
Fans, kW	5,190	3,160
Total, kW	8,730	3,835
Steam, 10^6 Btu/hr	96.6	89.4
Process water, 10^3 ℓ/hr (10^3 gal/hr)	122.3 (32.3)	111.7 (29.5)
Products		
Gypsum (10 wt % moisture), 10^3 kg/hr (tons/hr)	44.8 (49.4)	6.9 (7.6)
m^3/hr (yd^3/hr)	33.0 (43.1)	5.1 (6.6)
Incremental heat rate penalties		
Steam (basis gross heat rate), %	2.14	1.91
Electricity (basis unit net rating), %	1.75	0.77
Corrected net heat rate, Btu/kWh	9,931	10,168
Variation from base, %	-0.55	-1.21

Source: EPRI CS-1428

SO_2 Absorption – Hot flue gas discharges from the ID fans into a plenum from which it is distributed to the operating absorption trains. Before entering the absorbers the hot flue gas passes through low-pressure-drop spray tower prescrubbers where it is adiabatically cooled by sprays of recirculating fly ash slurry which also remove residual fly ash and some SO_2, SO_3 and chlorides. A slip stream of the recirculating slurry is sent to the absorber vessels where it is neutralized.

This bleed stream is controlled to maintain the suspended solids level in the recirculating slurry at about 5%. Make-up water is added to the prescrubber vessels to offset evaporation and bleed losses.

The cooled, saturated gas enters a distribution plenum within the absorber vessels, and is sparged into a tank containing a slurry of limestone and gypsum. The slurry is moderately stirred both by mechanical agitators and by oxidizing air which enters through distribution headers near the bottom of the vessel. The sparged gas creates a froth layer above the liquid in which SO_2 absorption and sulfite oxidation take place.

The sulfates formed in the froth layer precipitate as gypsum crystals which grow as they circulate in the slurry bath until they reach sufficient size to settle to the bottom of the reaction vessel. Gypsum slurry is continually withdrawn from the bottom of the reaction vessel. A portion of this slurry is recycled to the reaction tank to provide seed crystals, while the balance is sent to centrifuges for dewatering.

The treated flue gas passes through chevron mist eliminators and reheaters before being discharged through the chimney.

Waste Slurry Dewatering and Storage – Gypsum slurry from the absorption section is centrifuged to produce a wet gypsum cake, which is stockpiled and subsequently trucked off-site for landfill disposal. Centrate liquor is recycled to the slurry preparation area to minimize makeup water requirements.

Development Status: Development of the CT-121 process was initiated in Japan for oil firing in 1975 in an effort to reduce the cost and complexity of the commercial CT-101 process. Tests were initiated at the bench and laboratory scale, and finally at the 0.354 m^3/sec (750 scfm) level. A conceptual design of a 59 m^3/sec (125,000 scfm) plant has been completed, and a 23 MW pilot plant at Gulf Power Company's coal-fired Scholz Power Station started in mid-1978 and is currently undergoing testing.

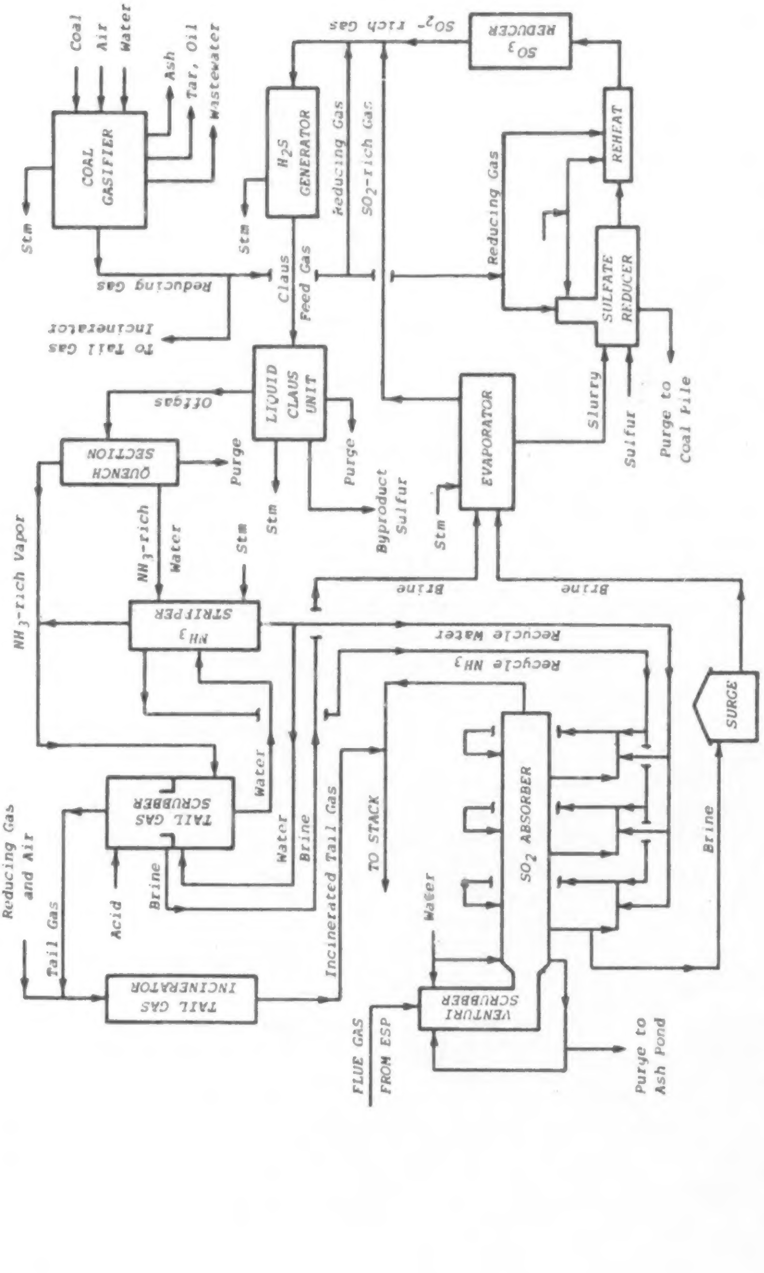
Ammonia Scrubbing Process

The ammonia scrubbing process proposed jointly by Catalytic, Inc. (a subsidiary of Air Products and Chemicals, Inc.) and Institut Francais du Petrole (IFP) uses an aqueous ammonia solution to absorb and remove SO_2 from flue gases. The resulting ammoniacal salt brine is then regenerated to yield elemental sulfur as an end product.

The SO_2 absorption section of the overall process involves ammonia scrubbing technology developed by Catalytic. The brine regeneration is based on technology developed and licensed by IFP.

Process Description: A schematic flow diagram of the Catalytic/IFP process is presented in Figure 3.3. Flue gas from an electrostatic precipitator (ESP) in a coal-burning power plant is first washed with water in a venturi scrubber to condition the flue gas and to remove chlorides, some sulfur trioxide and residual fly ash. The flue gas then enters the absorber.

Figure 3.3: Catalytic/IFP Ammonia Scrubbing Process


Coal
Air
Water
COAL GASIFIER
Ash
Tar, Oil
Wastewater
Stm
Reducing Gas
To Tail Gas Incinerator
Claus Feed Gas
H_2S GENERATOR
Stm
Reducing Gas
SO_2-rich Gas
SO_2- rich Gas
SO_3 REDUCER
REHEAT
Reducing Gas
SULFATE REDUCER
Slurry
Sulfur
Purge to Coal Pile
LIQUID CLAUS UNIT
Offgas
Purge
Stm
Byproduct Sulfur
EVAPORATOR
Stm
QUENCH SECTION
Purge
NH_3-rich Water
Stm
NH_3 STRIPPER
NH_3-rich Vapor
Brine
Brine
Recycle Water
Recycle NH_3
SURGE
Reducing Gas and Air
Tail Gas
TAIL GAS INCINERATOR
Acid
Brine
TAIL GAS SCRUBBER
Water
Water
Brine
Incinerated Tail Gas
TO STACK
Water
SO_2 ABSORBER
Brine
FLUE GAS FROM ESP
VENTURI SCRUBBER
Purge to Ash Pond

Source: EPRI CS-1381

Absorption Section – The Catalytic/IFP process uses multiple scrubbing stages in a horizontal absorber. The SO_2 absorption stages contain Koch Flexipac packing and the final stage is a spray chamber in which the flue gas is scrubbed with water to remove and recover any residual ammonia. Mist eliminators are included after the venturi and after each stage. Each stage of the absorber includes a liquid storage sump and circulation pump, and the circulation of the scrubbing liquid within each stage is controlled so as to achieve the desired ratio of liquid flow to gas flow. The scrubbing liquid cascades stage-by-stage to the first SO_2 absorption stage. In effect, the flow of the scrubbing liquid through the absorption stages is countercurrent relative to the flue gas flow, except that ammonia solution is fed to the individual stages as required to control the scrubbing liquor concentration.

SO_2 is removed from the flue gas by absorption in the ammonia solution which the gas contacts as it flows through the absorption stages of the absorber. The primary reactions occurring in the SO_2 absorption stages are:

$$SO_2 + 2NH_4OH \rightleftharpoons (NH_4)_2SO_3 + H_2O$$
$$SO_2 + (NH_4)_2SO_3 + H_2O \rightleftharpoons 2NH_4HSO_3$$
$$SO_3 + (NH_4)_2SO_3 \rightleftharpoons (NH_4)_2SO_4 + SO_2$$
$$2(NH_4)_2SO_3 + O_2 \rightleftharpoons 2(NH_4)_2SO_4$$

Only about 10% of the sulfur values are oxidized to $(NH_4)_2SO_4$ in the scrubber.

The flue gas is then scrubbed with water to remove any residual ammonia and passes through the final mist eliminator before leaving the absorber.

The scrubbing liquid leaving the absorber is a concentrated brine of these ammoniacal salts: ammonium sulfite, ammonium bisulfite and ammonium sulfate.

Regeneration Section (Reduction of Ammoniacal Salts) – The concentrated brine from the absorption section is sent to an intermediate surge storage tank from which the brine is pumped into a steam-heated forced circulation evaporator. Approximately 60% of the brine is evaporated and thermally decomposed to yield an SO_2-rich gas which also contains ammonia and water vapor.

The reactions occurring in the evaporator are:

$$(NH_4)_2SO_3 \rightarrow 2NH_3 + H_2O + SO_2$$
$$NH_4HSO_3 \rightarrow NH_3 + H_2O + SO_2$$

Since the sulfates in the brine do not decompose in the evaporator, a concentrated sulfate slurry is withdrawn from the evaporator bottom and sent to a sulfate reducing reactor. In the presence of molten sulfur, a temperature of about 600° to 700°F and a reducing atmosphere, the sulfates are decomposed in the sulfate reducer to yield an SO_2-rich gas which also contains ammonia, water vapor and a small amount of SO_3. The reactions occurring are:

$$(NH_4)_2SO_4 \rightarrow NH_4HSO_4 + NH_3$$
$$2NH_4HSO_4 + S \rightarrow 3SO_2 + 2NH_3 + 2H_2O$$
$$(NH_4)_2SO_4 \rightarrow SO_3 + 2NH_3 + H_2O$$

Reaction heat and the reducing atmosphere for the sulfate reducer are provided by the submerged exhaust of gas combusted within the reactor.

The gas yielded from the sulfate reducer passes through an in-line fired reheater and into an SO_3 reactor where the reducing atmosphere catalytically converts the small amount of SO_3 in the gas to SO_2.

The SO_2-rich gas streams from the brine evaporator and from the SO_3 reducer are combined with a reducing gas containing hydrogen and carbon monoxide. The combined gases are then sent to a hydrogen sulfide (H_2S) generator.

Regeneration Section (H_2S Generator) – To utilize the Claus reaction for converting SO_2 into elemental sulfur, two-thirds of the SO_2 must first be converted to H_2S since the Claus reaction requires a feed gas with approximately a 2:1 molar ratio of H_2S to SO_2. To accomplish this, a portion of the SO_2 in the SO_2-rich gas is reacted with the carbon monoxide and/or hydrogen in any ratio in the reducing gas as follows:

$$SO_2 + 3CO + H_2O \rightarrow H_2S + 3CO_2$$

$$SO_2 + 3H_2 \rightarrow H_2S + 2H_2O$$

After conversion of the required portion of the SO_2 to H_2S, the gases are cooled in a waste heat boiler and sent to a liquid phase Claus reactor.

The reducing gas required for generating the H_2S is provided by a commercial air-blown coal gasifier. The gasifier also provides gas for: the submerged combuster supplying hot exhaust for use in the sulfate reducer, the in-line fired reheater ahead of the SO_3 reducer and the final tail gas incinerator. The gasifier includes facilities for separation and removal of gasifier bottom ash as well as for cleansing the reducing gas of fly ash, tar and oil, and wastewater.

Regeneration Section (Liquid Phase Claus Unit) – The liquid phase Claus unit consists of a vertical tower about 75% full of a polyethylene glycol (PEG) and sulfur solution. The conversion of H_2S and SO_2 to elemental sulfur occurs in the liquid phase via the following reaction:

$$2H_2S + SO_2 \rightarrow 3S + 2H_2O$$

Most of the molten sulfur end product from the Claus unit is sent to storage and a small percentage is recycled to the sulfate reducer.

Regeneration Section (Ammonia Recovery and Recycle) – The Claus unit off-gas is quenched with water to reduce its temperature and to remove sulfur vapor and PEG from the gas. The gas then goes through a water-cooled condenser and into a condenser flash drum. The uncondensed vapor from the flash drum (which contains ammonia) is sent to a tail gas scrubber. The NH_3-rich aqueous solution from the condenser flash drum is routed through a steam-reboiled NH_3 stripper from which the following streams are produced:

> An overhead, NH_3-rich vapor which is sent to the tail gas scrubber along with vapor from the quench condenser flash drum.
>
> An overhead solution of aqueous ammonia which is recycled for reuse in the flue gas SO_2 absorber.
>
> A bottoms stream of water, containing some ammonia, which is also recycled for reuse in the flue gas SO_2 absorber. A part of this water is used as scrubbing liquid in the tail gas scrubber.

The tail gas scrubber is a two-stage unit. In the lower section, water from the NH_3 stripper bottoms is used to absorb NH_3 from the NH_3-rich tail gases and the water recycles back to the NH_3 stripper. In the upper section, a sulfuric acid solution removes the last traces of NH_3 from the tail gases and the resulting ammonium sulfate brine is sent to the evaporator and the sulfate reducer.

The scrubbed tail gases are then incinerated and blended with the flue gas from the SO_2 absorber to provide about 20°F of flue gas reheat.

Purge Streams – A slip-stream of the wash water from the flue gas inlet venturi scrubber is neutralized with lime and then filtered. The filtrate is recycled for reuse in the venturi scrubber. The sludge from the filter is purged to the power plant's ash pond.

The sulfate reducer is purged on a periodic basis to discharge any buildup of metallic salts which may originate with fly ash carryover into the SO_2 absorber.

A "sulfate cream" is periodically skimmed from the PEG-sulfur interface in the Claus reaction tower. The skimmed cream is separated to recover PEG and sulfur for recycle to the Claus reactor. The sulfate solution which remains is then purged to the sulfate reducer.

A slip-stream from the Claus off-gas quenching water is similarly separated to recover PEG for recycle to the Claus reactor. The sulfur compounds which have accumulated and combined to form a thiosulfate solution are also purged to the sulfate reducer.

As discussed above, the coal gasifier which produces reducing gas also discharges ash, tar and oil, and wastewater which will require appropriate handling and disposal.

Conoco Process

The Consol Flue Gas Desulfurization Process, developed by Conoco Coal Development Company (a subsidiary of Continental Oil Company), uses an aqueous solution of potassium carbonate and other potassium salts to absorb and remove SO_2 from flue gases. The resulting potassium bisulfite solution is converted to a potassium thiosulfate solution which is then reduced with carbon monoxide to yield a regenerated solution of potassium carbonate and hydrogen sulfide gas. The hydrogen sulfide gas is concentrated and converted to elemental sulfur as an end product.

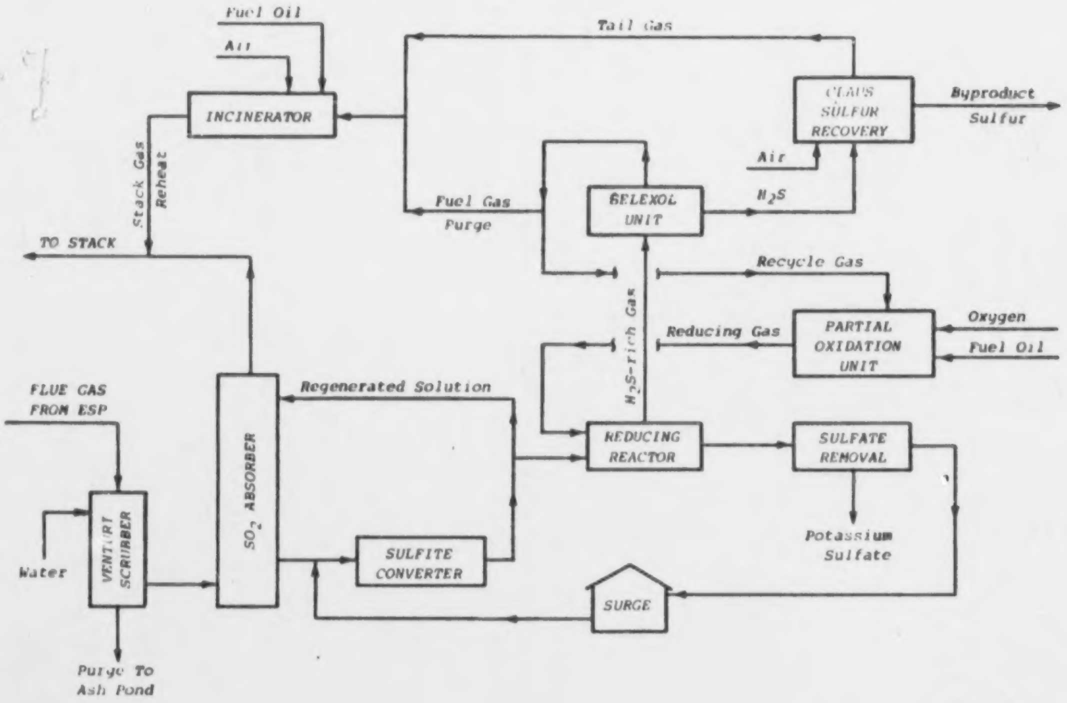
The Consol flue gas desulfurization process was offered by Conoco with The Rust Engineering Company participating as Conoco's engineering and construction subcontractor. The overall process includes two licensed subprocesses: the Selexol Process for concentrating hydrogen sulfide which is licensed by the Allied Chemical Company, and the Partial Oxidation Process for producing carbon monoxide (from heavy fuel oil) which is licensed by the Texaco Development Corporation.

Process Description: A schematic flow diagram of Conoco's process is presented in Figure 3.4. Flue gas from an electrostatic precipitator (ESP) in a coal-burning power plant is first washed with water in a venturi scrubber to remove chlorides, sulfur trioxide and residual fly ash. The flue gas then enters the SO_2 absorber.

Absorption Section – The flue gas flows upward through a packed SO_2 absorber where it is contacted by the downward flowing potassium carbonate solution which absorbs and removes SO_2 from the flue gas. The SO_2 reacts with the potassium carbonate (K_2CO_3) to form potassium bisulfite ($KHSO_3$) as shown below:

$$4SO_2 + 2K_2CO_3 + 2H_2O \rightarrow 4KHSO_3 + 2CO_2$$

Figure 3.4: Conoco Process


Fuel Oil
Air
INCINERATOR
Tail Gas
CLAUS SULFUR RECOVERY
Byproduct Sulfur
Air
Stack Gas Reheat
TO STACK
Fuel Gas Purge
SELEXOL UNIT
H_2S
Recycle Gas
H_2S-rich Gas
Reducing Gas
PARTIAL OXIDATION UNIT
Oxygen
Fuel Oil
FLUE GAS FROM ESP
SO_2 ABSORBER
Regenerated Solution
REDUCING REACTOR
SULFATE REMOVAL
Potassium Sulfate
Water
VENTURI SCRUBBER
SULFITE CONVERTER
SURGE
Purge To Ash Pond

Source: EPRI CS-1381

The absorbent solution leaves the absorber, is joined by regenerated solution returning to the absorption section and the combined stream passes through the sulfite converter. The regenerated solution contains almost equimolar amounts of potassium carbonate and potassium hydrosulfide (KHS). The KHS in the regenerated solution reacts with the $KHSO_3$ in the absorbent solution to yield potassium thiosulfate ($K_2S_2O_3$):

$$2KHS + 4KHSO_3 \rightarrow 3K_2S_2O_3 + 3H_2O$$

The potassium carbonate present in the sulfite converter does not enter into the reaction. The absorbent solution then recirculates from the sulfite converter back to top of the absorber for reuse. Thus, the absorbent solution entering the absorber is predominantly a potassium thiosulfate solution with a small amount of the active component (potassium carbonate) and a very small amount of potassium bisulfite.

Since potassium thiosulfate is a stable, nonoxidizable salt, the absorbent solution can be continuously recirculated through the SO_2 absorber. A small portion of the solution is withdrawn to be regenerated by reducing the potassium thiosulfate to potassium carbonate and potassium hydrosulfide so as to maintain the desired composition of the circulating absorbent solution.

Regeneration Section (Thiosulfate Reduction and Sulfate Removal) – The thiosulfate-rich solution from the absorption section is reacted with a carbon monoxide-rich reducing gas in a noncatalytic, thermal reductor. The reaction yields gaseous H_2S and regenerates the solution to potassium carbonate and potassium hydrosulfide:

$$3K_2S_2O_3 + 12CO + 5H_2O \rightarrow 2KHS + 2K_2CO_3 + 4H_2S + 10CO_2$$

The reaction occurs at about 450°F and 600 psig in an agitated reductor vessel.

After cooling and pressure letdown, the regenerated solution passes through a crystallizer and a filter. This removes any fly ash picked up in the absorber as well as a small amount of by-product potassium sulfate formed and accumulated in the absorbent solution. The regenerated solution then returns to the absorption section via an intermediate surge storage tank.

H_2S Concentration – The selexol process, licensed by the Allied Chemical Company, selectively absorbs H_2S from gases high in CO_2 content. The reductor off-gas entering the Selexol unit absorber contains about 11 and 48 volume percent of H_2S and CO_2, respectively. The H_2S is selectively absorbed in the lean Selexol absorbent (a solution of mixed polyethylene glycol dimethyl ethers) along with some CO_2. The remainder of the reductor off-gas passes through the absorber and exits the Selexol unit containing less than 125 ppm of H_2S.

The rich Selexol absorbent is then stripped to yield an off-gas which contains about 50 and 48 vol % of H_2S and CO_2, respectively. Thus, the H_2S has been separated from the reductor off-gas and has been concentrated by a factor of almost five.

The stripped Selexol absorbent is recycled to the Selexol absorber for reuse. The concentrated H_2S is sent to a conventional, vapor-phase Claus unit for conversion to elemental sulfur as an end product.

The sulfur-free reductor off-gas (less than 125 ppm of H_2S) from the Selexol absorber is divided into two streams. About 20% is burned as fuel gas for the incineration of the Claus unit tail gas. The remaining 80%, which is rich in CO_2, is recycled to the Partial Oxidation Unit for conversion of the CO_2 into CO-rich reducing gas.

Partial Oxidation Unit (Production of Reducing Gas) – The carbon monoxide-rich reducing gas, required for regenerating the potassium thiosulfate solution, is produced by the partial oxidation of fuel oil using the proprietary Texaco Partial Oxidation Process.

In the partial oxidation unit, heavy fuel oil is reacted with oxygen of 95% purity to produce a reducing gas containing carbon monoxide, carbon dioxide and hydrogen. After cooling in a waste heat boiler and water scrubbing for removal of carbon fines, the product reducing gas is sent to the potassium thiosulfate reductor.

Conditions in the partial oxidation unit are such that the reverse water gas reaction occurs:

$$CO_2 + H_2 \rightarrow CO + H_2O$$

Thus, the CO_2 recycled to the partial oxidation unit from the Selexol absorber off-gas acts to increase the yield of CO in the reducing gas from the partial oxidation unit. The flow of recycle CO_2 to the partial oxidation unit is controlled to maintain a 2:1 volume ratio of CO to CO_2 in the product reducing gas.

The carbon fines scrubbed from the product reducing gas are recycled to extinction within the partial oxidation unit.

The sulfur content of the fuel oil feedstock used in the partial oxidation unit is not a limiting constraint. Any H_2S produced in the reducing gas joins the H_2S produced in the potassium thiosulfate reductor and is subsequently converted to elemental sulfur in the Claus unit.

Claus Unit – The concentrated H_2S from the Selexol stripper is converted to elemental sulfur in a conventional, vapor-phase Claus unit. The overall reaction occurring in the Claus unit is:

$$2H_2S + O_2 \rightarrow 2S + 2H_2O$$

The Claus unit includes three converter stages to achieve a 95% conversion of H_2S to elemental sulfur.

The residual tail gas from the Claus unit is incinerated and blended with the flue gas from the SO_2 absorber to provide gas reheat.

Purge Streams – A slip-stream of the wash water from the flue gas inlet venturi scrubber is neutralized with lime and discharged to the power plant's ash pond.

The by-product potassium sulfate from the regeneration section will require appropriate handling and disposal.

Aqueous Carbonate Process

The Aqueous Carbonate Process, developed by Atomics International (since renamed the Environmental & Energy Systems Division of Rockwell International), contacts hot flue gases with an aqueous sodium carbonate solution in a spray dryer. The sodium carbonate reacts with and removes SO_2 from the flue gases, and the solution is evaporated to dryness by the hot flue gases. The dry reaction products (a mixture of sodium sulfite, sodium sulfate and unreacted sodium carbonate) are removed from the flue gases by passage through cyclones and an electrostatic precipitator. Subsequent processing of the reaction products yields regenerated sodium carbonate and hydrogen sulfide gas. The hydrogen sulfide gas is converted to elemental sulfur as an end product.

Process Description: A schematic flow diagram of Atomics International's Aqueous Carbonate Process is presented in Figure 3.5. Flue gas from the combustion air preheaters of the boilers in a coal-burning power plant first passes through cyclones for removal of about 80% of the fly ash. The flue gas then enters the spray dryer at a temperature of 250° to 350°F.

Reaction Section – The flue gas flows through a spray dryer where it is contacted in cross-flow with atomized droplets of an aqueous sodium carbonate solution. The reaction between the SO_2 in the flue gas and the atomized sodium carbonate solution (Na_2CO_3) yields sodium sulfite (Na_2SO_3) and sodium sulfate (Na_2SO_4):

$$Na_2CO_3 + SO_2 \rightarrow Na_2SO_3 + CO_2$$

$$Na_2CO_3 + SO_2 + 0.5O_2 \rightarrow Na_2SO_4 + CO_2$$

Any SO_3 in the flue gas also reacts with Na_2CO_3 to yield Na_2SO_4:

$$Na_2CO_3 + SO_3 \rightarrow Na_2SO_4 + CO_2$$

The sensible heat of the hot flue gases evaporates the reaction product sodium salts and unreacted sodium carbonate to dryness. The flue gas leaves the spray dryer at a temperature of 150° to 250°F, passes through cyclones and an electrostatic precipitator, and the cleansed flue gas then exits to the plant stack. Passage of the flue gas through the cyclones and precipitator removes the dry reaction products along with unreacted sodium carbonate and the residual fly ash. The mixture of fly ash and spent carbonate (i.e., sodium sulfite, sodium sulfate and unreacted sodium carbonate) collected in the cyclone and precipitator hoppers is pneumatically conveyed to a storage hopper. From there, it is fed to the regeneration section along with crushed coal from a coal storage hopper.

Regeneration Section (Sulfate Reduction) – The mixture of fly ash, spent carbonate and crushed coal enters the sulfate reducer which contains a pool of molten sodium salts and which operates at about 1800°F. At that temperature, the reductant carbon in the coal reacts with the sodium sulfate and sodium sulfite in the spent carbonate and reduces them to molten sodium sulfide as shown below:

$$2C + Na_2SO_4 \rightarrow Na_2S + 2CO_2$$

$$1.5C + Na_2SO_3 \rightarrow Na_2S + 1.5CO_2$$

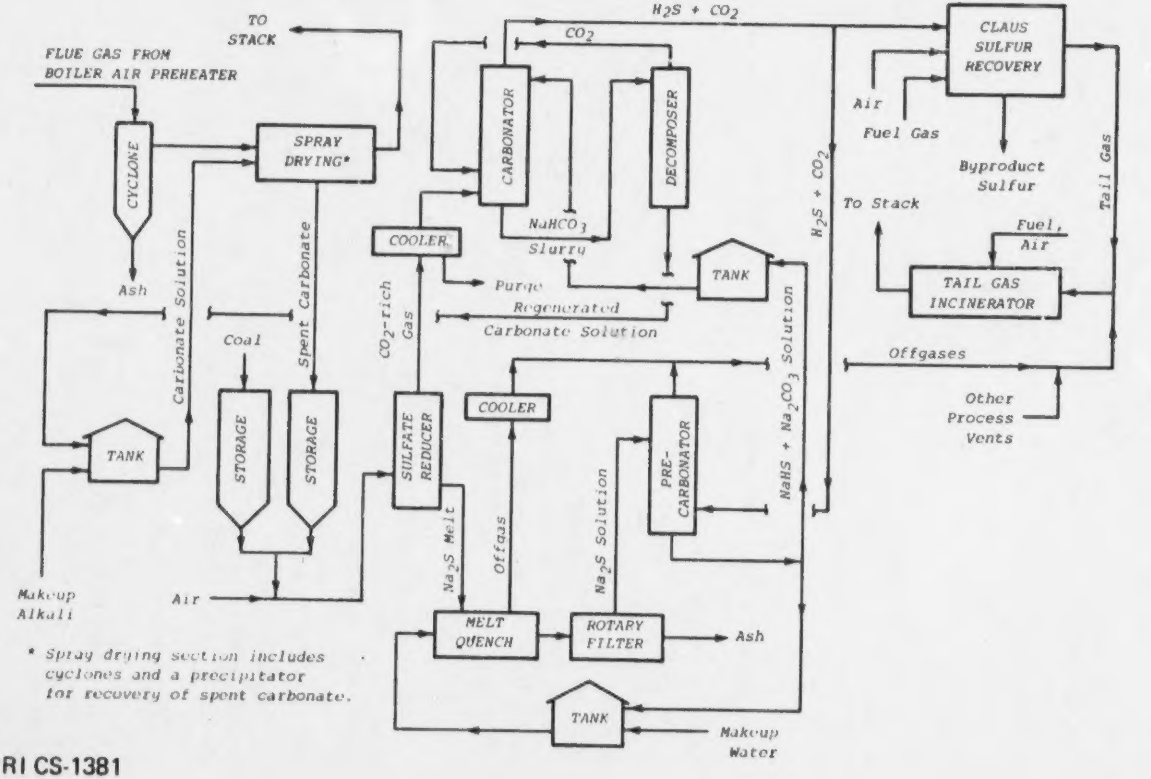
The unreacted sodium carbonate in the feed to the reducer does not enter into the reaction and goes through unchanged. The CO_2-rich gas yielded from the reducer is cooled by water-scrubbing and is sent to a carbonator in the carbonate recovery system.

The reducer melt, containing the molten Na_2S, is continuously discharged from the reducer. The melt is quenched, shattered and dissolved by recycle liquor. Off-gas vented from the quenching of the melt is cooled and sent to incineration. The Na_2S solution from the melt quench is filtered in a rotary vacuum filter which removes and discharges fly ash and any unreacted coal. The filtered solution is sent to a precarbonator tower where it is contacted with an H_2S and CO_2-rich gas derived from the carbonator in the carbonate recovery system. In the precarbonator, the Na_2S solution is converted to a solution of sodium hydrosulfide (NaHS) and sodium carbonate:

$$Na_2S + H_2S \rightarrow 2NaHS$$

$$2Na_2S + H_2O + CO_2 \rightarrow 2NaHS + Na_2CO_3$$

Figure 3.5: Atomics International Aqueous Carbonate Process


FLUE GAS FROM
BOILER AIR PREHEATER
CYCLONE
Ash
TO
STACK
SPRAY
DRYING*
Carbonate Solution
TANK
Makeup
Alkali
Coal
Spent Carbonate
STORAGE
STORAGE
Air
SULFATE
REDUCER
CO_2-rich
Gas
COOLER
Purge
CARBONATOR
$NaHCO_3$
Slurry
CO_2
DECOMPOSER
$H_2S + CO_2$
TANK
Regenerated
Carbonate Solution
Na_2S Melt
Offgas
COOLER
MELT
QUENCH
ROTARY
FILTER
Ash
Na_2S Solution
PRE-
CARBONATOR
$NaHS + Na_2CO_3$ Solution
TANK
Makeup
Water
$H_2S + CO_2$
CLAUS
SULFUR
RECOVERY
Air
Fuel Gas
Byproduct
Sulfur
Tail Gas
To Stack
Fuel,
Air
TAIL GAS
INCINERATOR
Offgases
Other
Process
Vents

* Spray drying section includes cyclones and a precipitator for recovery of spent carbonate.

Source: EPRI CS-1381

A portion of the precarbonated solution is recycled, via an intermediate surge storage tank, for reuse in quenching the reducer melt. The remainder of the stream (a solution of NaHS and Na_2CO_3) goes to the carbonator in the carbonate recovery system.

Off-gas from the precarbonator is sent to incineration along with the off-gas from the quenching of the reducer melt.

Regeneration Section (Carbonate Recovery) – The CO_2-rich gas from the sulfate reducer and the solution of NaHS and Na_2CO_3 from the precarbonator are contacted in the carbonate tower to produce a slurry of sodium bicarbonate:

$$NaHS + CO_2 + H_2O \rightarrow NaHCO_3 + H_2S$$

$$Na_2CO_3 + CO_2 + H_2O \rightarrow 2NaHCO_3$$

The off-gas from the carbonator contains H_2S and CO_2. A portion of the gas is sent to the precarbonator. The remainder is sent to the Claus unit for conversion to elemental sulfur.

The sodium bicarbonate slurry from the bottom of the carbonator is sent to a decomposer which produces the regenerated sodium carbonate solution for reuse in the flue gas spray dryer. The off-gas from the decomposer is rich in CO_2 and is recycled to the carbonator.

Claus Unit – A conventional, vapor-phase Claus unit converts the H_2S in the carbonator off-gases into elemental sulfur as an end product. The overall reaction occurring in the Claus unit is:

$$2H_2S + O_2 \rightarrow 2S + 2H_2O$$

The residual tail gas from the Claus unit is incinerated along with the off-gases from the melt quench, the precarbonator and other process vents. If the hot flue gas from the electrostatic precipitator requires any reheat to compensate for the heat removed in the spray dryer, the incinerated tail gas can be blended with the exit flue gas.

Purge Streams – There are no liquid purge streams from the Aqueous Carbonate Process. The discharge of ash and unreacted coal, from the filtering of the precarbonator feed solution, will require appropriate handling and disposal.

Two Advanced Processes in Use in Japan

Nippon Kokan Ammonia Scrubbing Process: Nippon Kokan (NKK) recently completed two large FGD plants to treat flue gas from iron ore sintering machines to produce ammonium sulfate utilizing ammonia in coke oven gas (1). Flue gas (1,120,000 Nm^3/hr) is treated with ammonium sulfite liquor to remove over 95% of SO_2. The liquor discharged from the scrubbers (two in parallel) containing ammonium bisulfite is contacted with coke oven gas to absorb ammonia. A portion of the resulting ammonium sulfite liquor is oxidized to produce ammonium sulfate.

Hydrogen sulfide in the coke oven gas is removed prior to the ammonia absorption. By-products from the sulfide removal unit are also oxidized and converted to ammonium sulfate.

The flue gas leaving the scrubber is passed through wet electrostatic precipitators (eight in parallel) and heated to 110°C by two Ljungstrom-type heat exchangers installed in parallel. No plume at all is observed from the stack.

Nippon Kokan has a similar plant at Fukuyama, with a capacity of treating 760,000 Nm^3/hr of flue gas. At this plant, the treated flue gas is heated by conventional oil firing using neither wet electrostatic precipitator nor heat exchanger. An appreciable plume is observed at the stack.

Electron Beam Process: Flue gas at about 100°C is mixed with NH_3 and exposed to electron beam radiation (2). About 80% of both SO_x and NO_x are removed with 2 Mrad of the beam forming fine crystals of ammonium nitrate sulfate double salt which are caught by an electrostatic precipitator for fertilizer use. Tests are in progress at a pilot plant (1 MW equivalent) of Nippon Steel.

Pircon-Peck Process

The Pircon-Peck flue gas desulfurization process was developed by Mr. Ladd Pircon and Dr. Ralph Peck. The process utilizes patented heterogeneous reactor technology developed by Mr. Pircon as cooling and absorber towers in conjunction with chemistry developed and demonstrated by Mr. Pircon and Dr. Peck at the Illinois Institute of Technology. The process is unique in that in addition to providing a means of controlling SO_2 emissions, it provides the owner with an opportunity to earn a profit on the invested capital.

Process Description: The Pircon-Peck FGD process, as shown in Figure 3.6a, utilizes "activated" phosphate rock, ammonia, and flue gas as raw materials to produce ammoniated phosphate fertilizers. Laboratory and pilot plant testing indicate that in the process of producing fertilizer, sufficient SO_2 can be removed from the flue gas to provide compliance with all EPA regulations, both existing and proposed.

The chemistry utilized to produce the product material can be summarized by three chemical reactions. These are as follows:

$$SO_2 + (NH_4)_2HPO_4 \xrightarrow{+ H_2O} NH_4HSO_3 + NH_4H_2PO_4$$

$$Ca_3(PO_4)_2 + 3NH_4HSO_3 \rightarrow 3CaSO_3 + (NH_4)_2HPO_4 + NH_4H_2PO_4$$

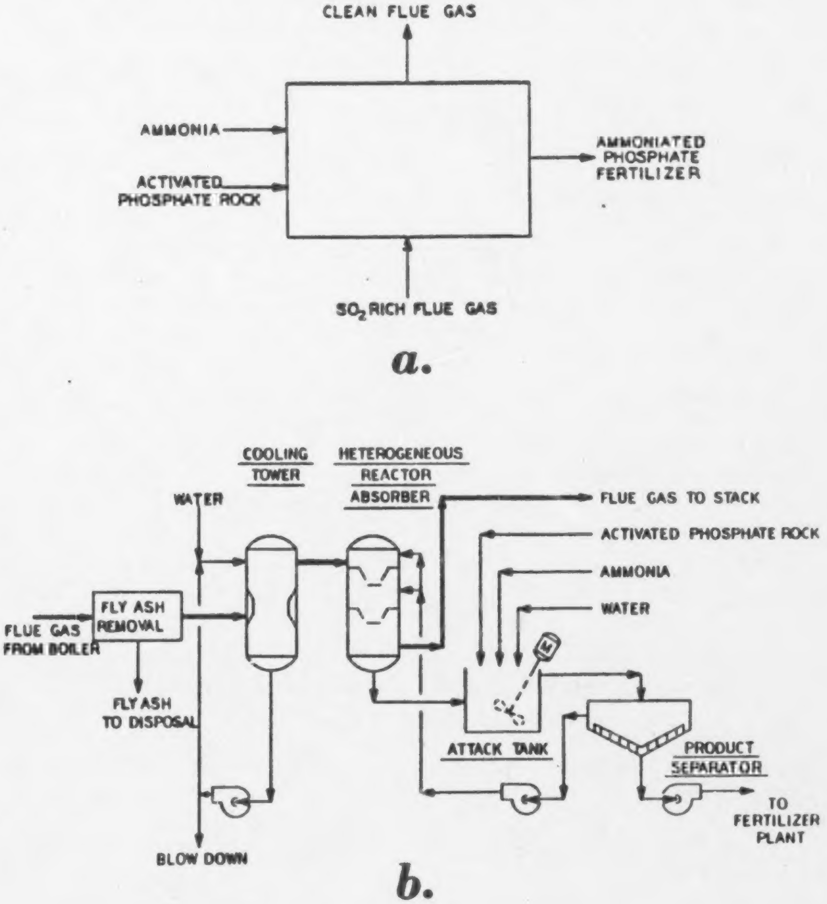
$$NH_4H_2PO_4 + NH_3 \rightarrow (NH_4)_2HPO_4$$

An important feature of this chemistry is the pretreatment of the phosphate rock which facilitates the reaction of the rock with the weak sulfurous acid produced in the scrubber. Standard phosphate fertilizer chemistry requires concentrated sulfuric acid to acidulate the rock and liberate the desired phosphate molecule.

In many ways, the Pircon-Peck process is similar to standard double alkali technology. In each system, a soluble alkali is utilized in the scrubber portion of the process and in each system this alkali is regenerated by a calcium source. In the Pircon-Peck process, the soluble alkali is diammonium phosphate and the calcium source is phosphate rock. However, where alkali regeneration with lime produces water as a by-product, diammonium phosphate regeneration with phosphate rock produces additional phosphate. Neutralization of this by-product with ammonia produces the final diammonium phosphate product.

In addition to producing diammonium phosphate, the regeneration reaction also produces calcium sulfite and sulfate. This material can be included in the fertilizer product or separated for disposal. If included in the fertilizer product, the material would be similar to superphosphate fertilizers presently produced by the fertilizer industry. These fertilizers contain the calcium sulfate solids produced by the acidulation of phosphate rock with sulfuric acid. While the final product form will be dictated by the local market conditions, the advantage of including the calcium sulfite/sulfate solids in the product is obvious.

Figure 3.6: Pircon-Peck Process


CLEAN FLUE GAS
AMMONIA
ACTIVATED PHOSPHATE ROCK
AMMONIATED PHOSPHATE FERTILIZER
SO_2 RICH FLUE GAS
a.
COOLING TOWER
HETEROGENEOUS REACTOR ABSORBER
WATER
FLUE GAS TO STACK
ACTIVATED PHOSPHATE ROCK
AMMONIA
WATER
FLY ASH REMOVAL
FLUE GAS FROM BOILER
FLY ASH TO DISPOSAL
ATTACK TANK
PRODUCT SEPARATOR
TO FERTILIZER PLANT
BLOW DOWN
b.

Source: Symposium–Vol 2

Figure 3.6b provides a view of the scrubber portion of the process. Flue gas enters the system and is first passed through a high efficiency particulate control device. The use of a particulate removal system minimizes the possibility of product contamination from fly ash. After fly ash removal, the gas is sent to a cooling tower where it is quenched and saturated to its adiabatic saturation temperature. The cooled and saturated flue gas is next processed through a heterogeneous reactor tower where the flue gas is contacted with a saturated solution of diammonium phosphate and the SO_2 is removed. The SO_2 free flue gas, after passing through a demister, is sent to the stack. If necessary, reheat can be added to aid in plume dispersion.

The SO_2-rich absorption liquor is taken from the heterogeneous reactor system and sent to an agitated attack tank where it is contacted with ammonia and phosphate rock. The slurry generated in the attack tank is next sent to a product separator (clarifier) where solid diammonium phosphate and calcium sulfite/sulfate crystals are withdrawn as a slurry. The product slurry is sent to a fertilizer plant where the product is converted to its final form. The overflow from the product separator is a saturated solution of diammonium phosphate and is sent back to the heterogeneous reactor tower for further reaction with SO_2.

The material from the product separator, on a dry basis, has an approximate analysis of 7-20-0 (N-P_2O_5-K_2O). This analysis can be adjusted as desired by the addition of ammonia, phosphoric acid, or other N-P-K materials.

As indicated, it is possible to produce product in various forms. Some alternates to granules with N and P_2O_5 are: pure diammonium phosphate, granules with various N, P and potassium values, suspensions, and flakes.

References

(1) Ando, J., SO_2 Abatement for Stationary Sources in Japan, EPA-600/7-78-210, November 1978. U.S. EPA.

(2) Japan Steel Federation, Status of Development of NO_x Removal Technology by Steel Industry (April 1978) (in Japanese).

COMPARISON OF ENERGY REQUIREMENTS OF SELECTED FGD SYSTEMS

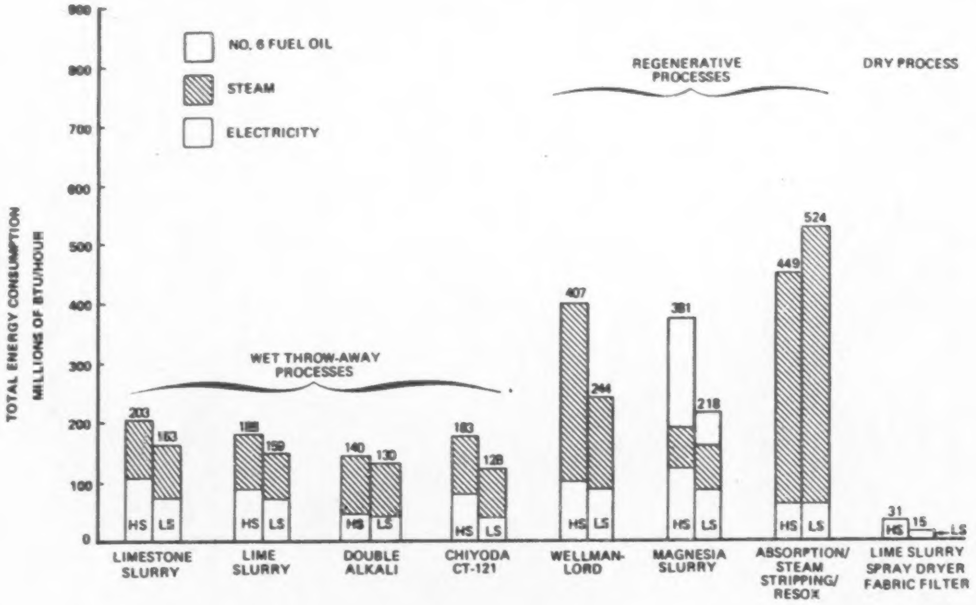
Energy Consumption: Figure 3.7 presents a comparison of the energy consumptions of FGD systems in terms of millions of Btu per hour for the high- and low-sulfur cases evaluated. Electrical energy consumption is represented in this figure by the equivalent heat input to the boiler required to produce this amount of energy. For purposes of comparison, the heat input to the model magnesia slurry systems provided by fuel oil has beeen treated as though an equivalent heat input were provided by steam. While the two energy sources may not be directly comparable on a per-Btu basis because of the higher temperatures attainable with fuel oil, the total heat input approach provides an indication of the overall energy consumption of each process for each type of coal.

Analysis of the individual energy inputs for each wet throw-away process and coal case indicates that the bulk of the electrical energy for all cases goes for fan power and absorbent pumping, both of which are related to gas flow volume. The other energy input, steam, is used entirely for flue gas reheat in all wet throw-away cases. Because the flue gas densities and reheat temperature differentials for both the high- and low-sulfur cases are comparable, this energy input is also primarily a function of gas volume.

Thus, it may be concluded that, for wet throw-away FGD processes employing flue gas reheat, total energy consumption is determined primarily by gas volume, and is only weakly influenced by total SO_2 removal requirements. Low-sulfur applications are therefore relatively more energy-intensive than high-sulfur applications for a given generating unit size and heat rate.

The general characteristic of the wet throw-away processes examined indicates that the total excess combustion air and total air in-leakage in the boiler draft system can have a significant effect on actual total FGD energy consumption. It also appears that there may be a basic level of energy consumption associated with the creation of the turbulent environment required for the absorption

Figure 3.7: Flue Gas Desulfurization—Total Energy Consumption



BASIS: NEW 500 MW UNIT, MAXIMUM LOAD, AVERAGE COAL SULFUR
HIGH SULFUR COAL: 4% S, 85% SO_2 REMOVAL
LOW SULFUR COAL: .48% S, 83% SO_2 REMOVAL
ELECTRICTY: 9,800-11,000 BTU/KWH
NO.6 FUEL OIL: HHV = 150,000 BTU/GAL.

HS-HIGH SULFUR COAL
LS-LOW SULFUR COAL

Source: EPRI CS-1428

of SO_2 by slurries of calcium compounds. Comparison of the total energy consumptions of the Chiyoda 121 and limestone slurry processes suggests that the Chiyoda 121 process absorption vessel design and process chemistry reduce the energy input required for absorption when using limestone slurry as the absorbent. This observation must, however, be considered tentative due to the relatively immature state of development of the Chiyoda process.

The total energy consumptions for all regenerative process cases are substantially higher than for the throw-away process cases due to the energy required to regenerate and reduce the absorbed SO_2. However, the requirements of the high- and low-sulfur cases still differ by less than a factor of two with the heat demands of the regeneration and reduction sections accounting for the bulk of both the total consumptions and the difference for each process.

In contrast to all of the other processes examined, the total energy consumption of the low-sulfur absorption steam stripping case is higher than that of the high-sulfur case. This difference is due to the fact that the specific steam consumption (pounds of steam per pound of SO_2 regenerated) of the stripping process increases as the sulfur loading of the absorbent liquor decreases.

The lime slurry spray dryer/fabric filter process employs no flue gas reheat and requires the lowest total energy input of all of the FGD processes examined.

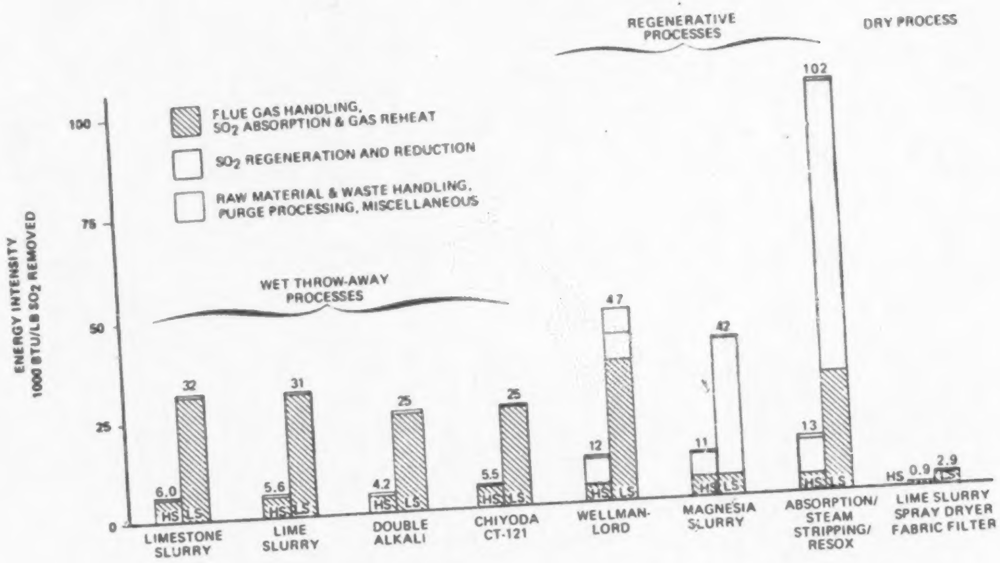
Energy Intensity: Figure 3.8 presents a comparison of the energy intensity of FGD systems in terms of thousands of Btus expended to remove one pound of SO_2 from the flue gas of the model generating units. Comparison of Figure 3.8 with Figure 3.7 indicates that for the four wet throw-away processes the absolute energy consumptions for the high- and low-sulfur cases differ by less than a factor of two, but that the energy intensities differ by approximately a factor of five with the low-sulfur case requiring the higher energy input per pound of SO_2 removed. Low-sulfur applications are therefore relatively more energy-intensive than high-sulfur applications for a given generating unit size and heat rate.

Comparison of the energy intensity characteristics of the three regenerative processes examined shows the same general trend found in the throw-away processes. The Wellman-Lord and magnesia slurry processes show somewhat reduced differences between the energy intensities for the high- and low-sulfur cases as compared with the wet throw-away processes. This is due largely to the fact that the energy demands of the regeneration and reduction sections are determined primarily by SO_2 removal rate rather than flue gas flow rate.

The absorption/steam stripping process shows a different trend from that of the other two regenerative processes because of the high steam consumption associated with stripping the relatively dilute absorber effluent liquor of the low-sulfur case. Further development of this process may reduce the specific steam consumption, and therefore the total energy intensity for all applications.

As would be expected, the spray dryer/fabric filter cases show the lowest total energy intensities of all of the FGD processes examined.

Figure 3.8: Flue Gas Desulfurization—Energy Intensity



BASIS: NEW 500 MW UNIT, MAXIMUM LOAD, AVERAGE COAL SULFUR
HIGH SULFUR COAL: 4% S, 85% SO_2 REMOVAL
LOW SULFUR COAL: .48% S, 83% SO_2 REMOVAL

HS-HIGH SULFUR COAL
LS-LOW SULFUR COAL

Source: EPRI CS-1428

An important consideration in developing a control process is complexity, its ability to adapt to varying coal grades, sulfur contents, and other variables. This Noyes Data report provides a process description, energy requirements, by-product management, and costs for each of nine flue gas desulfurization processes.

TECHNICAL FEASIBILITY AND ECONOMIC EVALUATION OF NINE PROCESSES

The material in this chapter is based on *NATO-CCMS Flue Gas Desulfurization Pilot Study—Phase II—Applicability Study,* August 1980, EPA-600/7-80-142, prepared by R.L. Torstrick, S.V. Tomlinson, J.R. Byrd and J.D. Veitch of Tennessee Valley Authority, Division of Energy Demonstrations and Technology, Office of Power, and R.W. Gerstle of PEDCo Environmental, Inc.

INTRODUCTION

The Phase I reports on 12 flue gas desulfurization processes for the North Atlantic Treaty Organization Committee on the Challenges of Modern Society (NATO-CCMS) FGD Study Group were reviewed by the NATO-CCMS delegates in April 1978 and nine of the processes were selected for comparative economic evaluations as Phase II of the study. The purpose of the Phase II study is to provide procedures and technical and economic data for the selection of FGD processes for specific applications. The study consists of technical feasibility and economic evaluations developed by the U.S. Tennessee Valley Authority (TVA) and a decision-chart selection procedure developed by PEDCo Environmental, Inc.

The basis of the economic evaluations is an FGD system for a new midwestern U.S. 500-MW power plant. The FGD system is designed for removal of 90% of the SO_2 in the flue gas. Six fuels, consisting of bituminous coals, lignite, and oil, are evaluated. Projection of costs for other conditions such as different power plant sizes and fuels with different sulfur levels are also provided.

The design and economic bases of this study differ from the Phase I study because of updated information and standardizations to provide comparable results between processes. It should be recognized that the data are based on U.S. conditions and that simplifying assumptions are made in the design model. These factors must be borne in mind when using the cost comparisons. Site-specific conditions may substantially alter the cost relationships of the processes.

The decision-chart system consists of an elimination procedure which rates the applicability of the processes relating the FGD process characteristics to specific site conditions. The system is an initial selection procedure which allows FGD selection efforts to be focused on the most promising processes.

PREMISES FOR THE ECONOMIC EVALUATIONS

The design and economic premises used were developed by TVA and others for comparative FGD cost studies using representative U.S. power plant conditions. The base case is a frontal-fired, balanced-draft 500-MW power plant constructed in the period 1977 to 1980 in the U.S. Midwest for startup in 1980. The plant is assumed to have a 30-year lifetime of 117,500 operating hours and to operate 6,000 hours in the first year. Fly ash removal of 99.2% by electrostatic precipitators (ESP) upstream of the FGD system is not included in FGD costs. The base case fuel is a 3.5% sulfur, 16% ash, 5,830 kcal/kg high heat rate bituminous coal. For this fuel and all other solid fuels it is assumed that 95% of the sulfur and 80% of the ash are emitted in the flue gas. For oil all of the sulfur in the fuel is assumed emitted in the flue gas.

The FGD systems are assumed to be installed downstream from the ESP units. For the wet systems the flue gas is supplied from a common plenum to four parallel trains of SO_x removal equipment including booster FD fans and reheat provisions. Presaturators to cool the gas from 149° to 53°C and mist eliminators to reduce liquid entrainment to 0.1% are provided. Indirect steam reheat (or direct-fired or oil-fired units used in the FGD process) is provided to reheat the gas entering the stack plenum to 79°C. The dry adsorption process is similar except two trains are used and reheat is not required. Costs for chloride removal facilities are included for the regenerable processes; however, costs for chloride disposal facilities are excluded. The FGD system consists of the common plenum and all equipment downstream to the stack plenum, including all raw material and effluent processing equipment and land requirements. Removal efficiencies are assumed to be 90% of the SO_2, 50% of the SO_3, 95% of the chloride, and 75% of the remaining fly ash in the flue gas.

The economic premises are based on U.S. regulated utility economics and financing. The costs estimated consist of capital costs for construction of the FGD system and annual revenue requirements for the first-year operation. All costs are based on midwestern U.S. costs using mid-1979 as the basis of capital costs and mid-1980 as the basis for annual revenue requirements. Capital costs consist of all direct and indirect costs for equipment, land, materials, labor, fees, services, and other construction costs required to install the FGD system. Annual revenue requirements, based on a first-year, 6,000-hour operation, consist of all raw material, labor, utility, and other conversion costs and indirect costs such as capital charges, taxes, and overheads. By-product sales for processes producing elemental sulfur, H_2SO_4, gypsum and sodium sulfate are included as credits. For the gypsum-producing processes, however, costs for disposal of gypsum may be obtained where necessary by substituting a disposal charge for stacking or landfill of the gypsum in place of the credit. These costs are projected to range between $4 and $9 per ton of dry solids, depending on site-specific conditions.

Case Variations: Case variations, in which design assumptions are varied to determine effect on costs, are also evaluated. Except for the seawater process, 0.8, 1.4, and 2.0% sulfur bituminous coal, 0.5% sulfur lignite, and 2.5% sulfur oil fuels are included. The seawater process is not evaluated for the 2.0 and 3.5% sulfur coals because those coals would require additional seawater in addition to condenser water. Other case variations consist of 200-, 700-, and 1,000-MW power plant sizes for the limestone sludge process, reheat to 53°C for the seawater process, and a sulfur production case for the Wellman-Lord process.

PROCESSES EVALUATED AND ACCURACY OF RESULTS

The nine process evaluations are based on data from a number of sources. Because of differences in the stage of development and amount of information available, the accuracy ranges of the economic results differ. Stage of development is difficult to quantify and is not considered in these evaluations. The accuracy ranges can, however, be related to the amount of information available. Normally, actual investment costs may be expected to depart from those shown in the economic evaluation by a factor of 0.7 to 1.5. Because of the extensive information available for some of the processes evaluated, however, smaller ranges can be projected. The accuracy ranges of each process, based on the amount of information available, are shown in the tabulated results. The nine processes evaluated are discussed below.

Limestone Sludge Process

Process Description: The limestone slurry process is designed to use a mobile-bed absorber, with presaturator and mist eliminator. The mist eliminator is equipped for upstream and downstream wash with fresh makeup water. The flue gas from the common plenum is cooled in the presaturator, scrubbed with limestone slurry in the absorber, passed through the mist eliminator, and is reheated before being vented to the stack plenum.

Limestone slurry is recirculated through the absorber and an external surge tank. The slurry is maintained at 15% solids by withdrawal of a purge stream and addition of fresh slurry. The reaction of SO_2 with $CaCO_3$ in the limestone is assumed to produce 80% calcium sulfite hemihydrate ($CaSO_3 \cdot \frac{1}{2}H_2O$) and 20% gypsum ($CaSO_4 \cdot 2H_2O$). The purge stream containing these salts, unreacted limestone, and minor impurities is pumped one mile to an earthen-diked, clay-lined pond where it settles to a sludge of about 40% solids. Supernate water is returned from the pond for reuse in the process.

The feed preparation area consists of two trains of crushers and wet ball mills serving all four absorber trains. As-received 40 mm maximum-size crushed limestone is further reduced in crushers and processed in wet ball mills to a 60% solids slurry with a particle size of about 70% less than 0.003 mm that is metered to the scrubber slurry loop.

Specific Process Premises:

- The flue gas is assumed to be cooled from 149° to 53°C (300° to 127°F) in the presaturator at an L/G ratio of 0.5 ℓ/m^3 (4 $gal/10^3\ aft^3$).
- The absorber is a mobile-bed-type turbulent contact absorber with a flue gas superficial velocity of 3.8 m/sec (12.5 ft/sec) and a pressure drop, including the mist eliminator, of 2.14 kPa (8.6 inches H_2O). An L/G ratio of 6.7 ℓ/m^3 (50 $gal/10^3\ aft^3$).
- Stoichiometry is 1.4 mols of $CaCO_3$ to 1.0 mol of SO_2 removed and 1.0 mol of $CaCO_3$ to 2.0 mols of HCl removed.

Energy Requirements: For base-case conditions, reheat of the cleaned gas from 53° to 79°C requires 42.2×10^3 kg/hr (93,050 lb/hr) of 243°C (470°F) steam at 3.55×10^3 kPa absolute pressure (500 psig) equivalent to approximately 17.62×10^6 kcal/hr.

The electrical power demand for the base-case limestone sludge process is about 7,995 kW or 1.6% of the rated output of a 500-MW power plant. For 6,000 hours of operation the annual electrical energy consumption is 48.0 x 10^6 kWh.

The total equivalent energy consumption for the base case is approximately 37.7 x 10^6 kcal/hr or 3.3% of the input energy required for the 500-MW power unit.

By-Product Management: ESP units remove 99.2% of the fly ash from the flue gas and, therefore, only a small amount of fly ash is found in the FGD process sludge. (Fly ash emission from oil-fired units does not exceed the EPA particulate emission standard and fly ash collection facilities are not included in oil-fired power plant design.) Projected mass flow rates of wastes for the base case are shown below:

Component	Kilograms per Hour	Pounds per Hour
$CaSO_3 \cdot ½H_2O$	16,550	36,480
$CaSO_4 \cdot 2H_2O$	5,670	12,500
$CaCO_3$	6,448	14,215
$CaCl_2$	433	955
Mg	39	85
Fly ash	149	329
Inerts	1,236	2,726
Total	30,525	67,290

Based on a 30-year life for both the power unit and the FGD units, the sludge disposal pond for the base case requires approximately 123 ha (305 acres). It is designed for an optimum depth of approximately 6.1 m (20 ft).

Lime Sludge Process

Process Description: Lime slurry scrubbing differs from limestone scrubbing in this study only in the raw material used as SO_2 absorbent and in the method of preparing the scrubbing slurry.

Pebble lime is slaked in two parallel slakers at a slurry concentration of 60% solids and combined with scrubber effluent slurry and recycle pond water to control the concentration of the recirculating slurry at approximately 15% solids. The flue gas is cooled in a presaturation chamber and passed through a mobile-bed absorber. The lime slurry circulates through the absorber where it reacts with the SO_2 in the cooled flue gas. Mist eliminators equipped for upstream and downstream wash with fresh makeup water control entrainment carryover in the gas stream. A bleedstream from the recirculation tank is pumped to an earthen-diked, clay-lined pond one mile away where it settles to form a sludge containing approximately 40% solids. The sludge is assumed to be 80% $CaSO_3 \cdot ½H_2O$ and 20% gypsum. Pond supernate is recycled to the slakers and the absorber recirculation tank to maintain closed-loop operation. Scrubber outlet gas is reheated to 79°C (175°F) by indirect steam heat before entering the stack.

Specific Process Premises:

- The flue gas is assumed to be cooled from 149° to 53°C (300° to 127°F) in the presaturator at an L/G ratio of 0.5 ℓ/m^3 (4 gal/10^3 aft^3).
- The absorber is a mobile-bed type with a flue gas superficial velocity of 3.8 m/sec (12.5 ft/sec) and a pressure drop of 2.14 kPa (8.6 inches H_2O), including the mist eliminator. An L/G ratio of 7.4 ℓ/m^3 (55 gal/10^3 aft^3) is used.

- Stoichiometry is 1.05 mols of CaO to 1.0 mol of SO_2 removed and 1.0 mol of CaO to 2.0 mols of HCl removed.

Energy Requirements: For base-case conditions, reheat of the cleaned gas from 53° to 79°C requires 42.1 x 10^3 kg/hr (92,740 lb/hr) of 243°C (470°F) steam at 3.55 x 10^3 kPa absolute pressure (500 psig), equivalent to about 17.56 x 10^6 kcal/hr.

The electrical power demand for the base-case lime-sludge process is about 7,448 kW or 1.5% of the rated output of a 500-MW power plant. For 6,000 hours of operation, the annual electrical energy consumption is 44.7 x 10^6 kWh.

The total equivalent energy consumption for the base case is approximately 36.41 x 10^6 kcal/hr or 3.2% of the input energy required for the 500-MW power unit.

By-Product Management: Electrostatic precipitators remove 99.2% of the fly ash from the flue gas and, therefore, only a small amount of fly ash is found in the FGD process sludge. (Fly ash emission from oil-fired units does not exceed the EPA particulate emission standard and fly ash collection facilities are not included in oil-fired power plant design.) Projected mass flow rates of by-product wastes for the base case are shown below.

Component	Kilograms per Hour	Pounds per Hour
$CaSO_3 \cdot \frac{1}{2}H_2O$	16,550	36,480
$CaSO_4 \cdot 2H_2O$	5,670	12,500
$Ca(OH)_2$	451	995
$CaCl_2$	433	955
Mg	123	271
Fly ash	149	329
Inerts	102	225
Total	23,478	51,755

Based on a 30-year life for both the power unit and the FGD unit, the sludge disposal pond for the base case requires approximately 104 ha (256 acres). It is designed for an optimum depth of approximately 5.8 m (19 ft).

Double-Alkali Sludge Process

Process Description: The double-alkali process evaluated has been generalized from several concentrated-mode double-alkali processes in the United States. A two-tray tower absorber with presaturator and mist eliminator is used. The scrubbing liquor is a solution of sodium salts of which sodium sulfite (Na_2SO_3) is the major active component. The SO_2 reacts with the Na_2SO_3 to form sodium bisulfite ($NaHSO_3$). A bleedstream of absorber liquor is treated with slaked lime to precipitate calcium sulfur salts and regenerate the Na_2SO_3.

Pebble lime is slaked and then reacted with a bleedstream of absorber effluent in agitated tanks. The reaction product, predominately calcium sulfite, flows to a thickener where the slurry is concentrated to 40% solids. This stream is further dewatered using drum filters to produce a cake containing 55% solids. The filter is designed with two wash sections to minimize sodium loss. The filter cake is conveyed to a reslurry tank where it is mixed with pond return water to a 15% solids slurry. The slurry is pumped to an earthen-diked clay-lined pond one mile away where the solids in the slurry settle to form a sludge containing approximately 40% solids. Makeup soda ash is added to the regenerated scrubber liquor at the thickener overflow storage tank.

Specific Process Premises:

- The flue gas is cooled from 149° to 53°C (300° to 127°F) and saturated in the presaturator. The presaturator has an L/G ratio of 0.5 ℓ/m^3 (4 gal/10^3 aft^3).
- A two-tray tower absorber with a superficial velocity of 2.1 m/sec (7 ft/sec), and a pressure drop, including the mist eliminator, of 1.25 kPa (5 inches H_2O) is used. An L/G ratio of 0.5 ℓ/m^3 (4 gal/10^3 aft^3) is used for recycle liquor to the absorber and an L/G ratio of 0.4 ℓ/m^3 (3 gal/10^3 aft^3) is used for the regenerated scrubbing liquor to the absorber.
- Stoichiometry is 1.0 mol of CaO to 1.0 mol of SO_2 removed and 1 mol of CaO to 2 mols of HCl removed.
- Oxidation of 10% of the SO_2 removed to sulfate is assumed. The remainder is assumed to be in sulfite form.

Energy Requirements: For base-case conditions, reheat of the cleaned gas from 53° to 79°C requires 42.2 x 10^3 kg/hr (93,060 lb/hr) of 243°C (470°F) steam at 3.55 x 10^3 kPa absolute pressure (500 psig), equivalent to about 17.62 x 10^6 kcal/hr.

The electrical power demand for the base case is estimated to be about 3,981 kW or 0.8% of the rated output of a 500-MW power plant. For 6,000 hours of operation the annual electrical energy consumption is 23.9 x 10^6 kWh.

The total equivalent energy consumption for the base case is approximately 28.61 x 10^6 kcal/hr or 2.5% of the input energy required for the 500-MW power unit.

By-Product Management: ESP units remove 99.2% of the fly ash from the flue gas and, therefore, only a small amount of fly ash is found in the FGD process sludge. (Fly ash emission from oil-fired units does not exceed the EPA particulate emission standard and fly ash collection facilities are not included in oil-fired power plant design.) Projected mass flow rates of by-product wastes for the base case are shown below:

Component	Kilograms per Hour	Pounds per Hour
$CaSO_3 \cdot ½H_2O$	18,266	40,270
$CaSO_4 \cdot 2H_2O$	2,158	4,758
$Ca(OH)_2$	209	461
Na_2SO_3	319	703
Na_2SO_4	148	326
NaCl	456	1,006
Mg	117	258
Fly ash	149	329
Inerts	98	215
Total	21,920	48,326

Based on a 30-year life for both the power unit and the FGD unit, the sludge disposal pond for the base case requires approximately 99 ha (245 acres). It is designed for an optimum depth of approximately 5.8 m (19 ft).

Seawater Process

Process Description: The seawater process uses seawater from the power plant condensers as the scrubbing agent. Because the amount of condenser seawater is limited, the 3.5% and 2.0% sulfur coals are not included in the seawater process evaluation. The 149°C (300°F) flue gas is cooled to 53°C (127°F) in a presaturator and scrubbed in a countercurrent packed tower absorber to remove SO_2, SO_3, HCl, CO_2, and some residual fly ash. Seawater at 26°C (79°F) is used in both the presaturator and absorber in a one-pass flow. The flue gas is cooled to 27°C (81°F) in the SO_2 absorber.

Presaturator and absorber effluent at a pH of approximately 3.0 (based on SO_2 content only) is treated with additional condenser seawater to increase the pH to 6, treated with sparged air to oxidize 75% of the sulfite to sulfate, and returned to the sea. The chemical oxygen demand of the waste is estimated to be 3.1 mg O_2 per liter or less. Reheat of the flue gas to 79°C (175°F) is included in the process. A case variation of reheat to 53°C (127°F) is also included.

Specific Process Premises:

- A packed-bed absorber with a presaturator and mist eliminator is used. Pressure drop in the unit is 1.05 kPa (4.2 inches H_2O) and the superficial velocity in the absorber is 1.8 m/sec (6 ft/sec).
- The presaturator L/G ratio is 0.5 ℓ/m^3 (4 gal/10^3 aft^3) and the absorber L/G ratio is 8.0 ℓ/m^3 (60 gal/10^3 aft^3).
- The total alkalinity of the seawater is assumed to be 2.4 meq/ℓ as $CaCO_3$.
- Oxidation of SO_x to SO_4 is assumed to be 75% in the oxidation tank.
- Only condenser seawater from the power plant is used for neutralization; no alkali or additional seawater is added.

Energy Requirements: For base-case conditions, reheat of the cleaned gas from 27° to 79°C requires 68.7 x 10^3 kg/hr (151,500 lb/hr) of 243°C (470°F) steam at 3.55 x 10^3 kPa absolute pressure (500 psig) equivalent to about 28.70 x 10^6 kcal/hr.

The electrical power demand for the seawater process, base case, is estimated to be about 7,012 kW or 1.4% of the rated capacity of a 500-MW power plant. For 6,000 hours of operation, the annual electrical energy consumption is 42.1 x 10^6 kWh.

The total equivalent energy consumption for the base case is approximately 47.79 x 10^6 kcal/hr or 4.2% of the input energy required for the 500-MW power unit.

By-Product Management: ESP units remove 99.2% of the fly ash from the flue gas and, therefore, only a small amount of fly ash is found in the effluent which is pumped offshore for disposal. (Fly ash emission from oil-fired units does not exceed the EPA particulate emission standard and fly ash collection facilities are not included in oil-fired power plant design.) Projected mass flow rates of by-product wastes for the base case are shown below.

Component	Kilograms per Hour	Pounds per Hour
Total effluent	72,124,300	158,864,200
$SO_4^=$	4,040	8,900
$SO_3^=$	1,110	2,450
HCl	470	1,050
Fly ash	50	120

Approximate chemical oxygen demand (COD) for this effluent is estimated at 3.1 mg O_2 per liter.

Lime Gypsum (Saarberg-Hölter) Process

Process Description: The Saarberg-Hölter process removes SO_2 using a clear alkaline scrubbing solution in a Rotopart, a patented modular-design absorber. The flue gas enters the Rotopart vessel at 149°C (300°F) and is adiabatically cooled to 53°C (127°F) by contact with clear scrubbing solution. The flue gas and the scrubbing solution are cocurrently contacted in vertical scrubbing tubes of the Rotopart vessel. The scrubbing tubes have no packing but are equipped with injection nozzles and special shedding rings to promote liquid-gas contact. SO_2, SO_3, HCl, and some residual fly ash are removed from the gas stream. Gas and washing fluid are centrifugally separated in the separator section of the Rotopart vessel. No further demisting is required.

SO_2 in the flue gas reacts with $CaCl_2$ and $Ca(OH)_2$ in the scrubbing solution to form soluble $Ca(HSO_3)_2$ and HCl. The system uses formic acid as a buffering agent to maintain a pH of about 4.5. The acidic absorber effluent flows by gravity to an oxidizer vessel where air blown through the liquid oxidizes $Ca(HSO_3)_2$ to $CaSO_4 \cdot 2H_2O$ and H_2SO_4. $Ca(OH)_2$ is added at the oxidizer to neutralize the liquid and to replenish calcium in the solution. Makeup formic acid is also added at the oxidizer. Oxidizer effluent is pumped to a thickener where the suspension of the $CaSO_4 \cdot 2H_2O$ crystals is thickened to a 15% solids slurry. Slurry from the thickener is filtered to produce an 80% solids cake. The filter cake is conveyed to a storage area. Filtrate is recycled to the thickener and thickener overflow at a pH of 10.5 is returned to the absorber for use as scrubbing liquid.

Specific Process Premises:

- The Rotopart has a superficial velocity of 12.2 m/sec (40 ft/sec) in the scrubber tubes and a pressure drop through the unit of 2.44 kPa (9.8 inches H_2O). The L/G ratio is proprietary.
- The stoichiometry is 1.01 mols of CaO to 1.00 mol of SO_2 removed and 1.0 mol of CaO to 2.0 mols of HCl removed.
- Complete oxidation of SO_2 removed to sulfate in the form of gypsum is assumed.

Energy Requirements: For base-case conditions, reheat of the cleaned gas from 53° to 79°C requires 42.4 x 10^3 kg/hr (93,460 lb/hr) of 243°C (470°F) steam at 3,550 kPa absolute pressure (500 psig), equivalent to about 17.7 x 10^6 kcal/hr.

The electrical power demand for the base-case lime gypsum process is about 9,701 kW or 1.9% of the rated output of a 500-MW power plant. For 6,000 hours of operation, the annual electrical energy consumption is 58.2 x 10^6 kWh.

The total equivalent energy consumption for the base case is approximately 39.70 x 10^6 kcal/hr or 3.7% of the input energy required for the 500-MW power unit.

By-Product Management: Electrostatic precipitators remove 99.2% of the fly ash from the flue gas and, therefore, only a small amount of fly ash is found in the FGD process by-product. (Fly ash emission from oil-fired units does not exceed the EPA particulate emission standard and fly ash collection facilities are not included in oil-fired power plant design.) Projected mass flow rates of by-product for the base case are shown below.

Component	Kilograms per Hour	Pounds per Hour
$CaSO_4 \cdot 2H_2O$	27,640	60,880
$CaSO_3 \cdot \frac{1}{2}H_2O$	82	180
$CaCl_2$	440	968
$Mg(OH)_2$	295	650
Fly ash	149	329
Inerts	102	225
Total	28,708	63,232

The process is evaluated on the basis of 30-day storage of gypsum by-product. A 0.4 ha (1 acre) storage area has been provided for the base case and all fuel variations.

Jet-Bubbling Limestone (Chiyoda Thoroughbred 121) Process

Process Description: The Chiyoda Thoroughbred 121 process is a forced-oxidation limestone-scrubbing process which produces gypsum. It was developed from Chiyoda's dilute sulfuric acid process, the Thoroughbred 101.

Absorption, oxidation, and crystallization are accomplished in the same reactor vessel. The flue gas is cooled in a presaturator chamber and fed to the agitated jet-bubbling reactor. Air and a limestone slurry of 15% solids are introduced to the reactor where SO_2 is absorbed from the flue gas, oxidized to SO_4, and reacted with the limestone to form gypsum (calcium sulfate dihydrate). A bleedstream containing the gypsum crystals is pumped to a thickener. The thickener underflow containing 40% solids is filtered to approximately 80% solids. The filter cake is conveyed to a storage area and the filtrate is returned to the system for use in the wet ball mills.

Specific Process Premises:

- The reactor for absorption, oxidation, and neutralization is a Chiyoda jet-bubbling reactor equipped with a presaturator and mist eliminator. Pressure drop is 3.94 kPa (15.8 inches H_2O). Complete oxidation of absorber SO_2 to gypsum is assumed.
- A stoichiometry of 1.0 mol of $CaCO_3$ to 1.0 mol SO_2 removed and 1.0 mol of $CaCO_3$ to 2.0 mols of HCl removed is used.

Energy Requirements: For base-case conditions, reheat of the cleaned gas from 53° to 79°C requires 43.5 x 10^3 kg/hr (95,900 lb/hr) of 243°C (470°F) steam at 3.55 x 10^3 kPa absolute pressure (500 psig) equivalent to about 18.16 x 10^6 kcal/hr.

The electrical power demand for the Chiyoda Thoroughbred 121 process, base case, is estimated to be about 8,090 kW or 1.6% of the rated production of a 500-MW power plant. For 6,000 hours of operation, the annual electrical energy consumption is 48.5 x 10^6 kWh.

The total equivalent energy consumption for the base case is approximately 38.53 x 10^6 kcal/hr or 3.4% of the input energy required for the 500-MW power unit.

By-Product Management: ESP units remove 99.2% of the fly ash from the flue gas and, therefore, only a small amount of fly ash is found in the FGD process by-product. (Fly ash emission from oil-fired units does not exceed the

EPA particulate emission standard and fly ash collection facilities are not included in oil-fired power plant design.) Projected mass flow rates of by-product for the base case are shown below.

Component	Kilograms per Hour	Pounds per Hour
$CaSO_4 \cdot 2H_2O$	27,751	61,180
$CaCl_2$	456	1,005
Mg	27	60
Fly ash	149	329
Inerts	884	1,948
Total	29,267	64,522

The process is evaluated on the basis of 30-day storage of saleable gypsum by-product. A 0.4 ha (1 acre) storage area has been provided for base case and all fuel variation estimates.

Magnesium Oxide Process

Process Description: The magnesium oxide process is an absorbent-regenerating process which produces sulfuric acid. The flue gas is scrubbed with an MgO slurry, producing $MgSO_3$ and $MgSO_4$, which are dried and calcined to regenerate SO_2 and MgO. The SO_2 is processed to sulfuric acid and the MgO is returned to the system. A spray grid absorber with a chloride scrubber and mist eliminator is used. Makeup and regenerated MgO are slurried into a bleedstream of liquor from the absorber and recycled to the absorber. Humidification losses are added as a fresh water upstream wash for the mist eliminator.

Effluent from the absorber, containing approximately 15% solids, is pumped to two parallel centrifuges to separate the solids from the liquor. The centrate is returned to the absorber system. The centrifuge cake is dried in an oil-fired rotary dryer. The dryer off-gas is cleaned in a cyclone and fabric dust collector. A portion of the gas is recycled to the dryer combustion chamber for temperature control and the remainder is routed to the absorbers.

The discharge from the dryer and filters is transferred to an oil-fired fluid-bed calciner which contains a single calcination bed operating at 871°C (1600°F). The calciner off-gas, containing SO_2, is partially cleaned in a cyclone, cooled to about 371°C (700°F) in a waste heat boiler, and fed to a fabric filter for final cleaning before entering the sulfuric acid unit. The MgO collected in the cyclone and bag filter is returned to the absorber feed preparation system.

A complete 390 metric ton/day contact sulfuric acid plant is provided for production of 98% acid from the SO_2 gas. The sulfuric acid is stored in 30-day capacity tanks. Tail gas from the acid plant is recycled to the absorber.

Specific Process Premises:

- The flue gas is assumed cooled from 149°C (300°F) to 53°C (127°F) and saturated in the presaturator chloride scrubber. The presaturator chloride scrubber L/G ratio is 0.5 ℓ/m^3 (4 gal/10^3 aft^3).
- A mobile-bed absorber with a superficial velocity of 3.8 m/sec (12.5 ft/sec) and a pressure drop, including the mist eliminator, of 1.99 kPa (8 inches H_2O). The absorber L/G ratio is 3 ℓ/m^3 (20 gal/10^3 aft^3).
- The stoichiometry is 1.05 mols of MgO to 1.0 mol of SO_2 removed.

Energy Requirements: For base-case conditions, reheat of the cleaned gas from 53° to 79°C requires 43.4 x 10^3 kg/hr (95,700 lb/hr) of 243°C (470°F) steam at 3.55 x 10^3 kPa absolute pressure (500 psig) equivalent to about 18.12 x 10^6 kcal/hr.

The electrical power demand for the base-case magnesium oxide process is about 9,109 kW or 1.8% of the rated output of a 500-MW power plant. For 6,000 hours' operation, annual electrical energy consumed is 54.65 x 10^6 kWh.

Fuel oil provides energy for the dryer and calciner. The total fuel oil consumption of approximately 3,690 ℓ/hr (975 gal/hr) is equivalent to 33.78 x 10^6 kcal/hr.

Waste heat is recovered in the acid production area and in the calcining area to produce 5.76 x 10^3 kg/hr (12,700 lb/hr) of 186°C (367°F) steam at 1.14 x 10^3 kPa absolute pressure (150 psig). This steam is equivalent to 3.6 x 10^6 kcal/hr and is considered a heat credit in the determination of total equivalent energy consumption.

The total equivalent energy consumption for the base case is approximately 70.97 Mkcal/hr or 6.26% of input energy required for the 500-MW power unit.

By-Product Management: ESP units remove 99.2% of the fly ash from the flue gas and, therefore, only a small amount of fly ash is found in the chloride scrubber effluent. (Fly ash emission from oil-fired units does not exceed the EPA particulate emission standard and fly ash collection facilities are not included for these units.) Projected mass flow rates of the by-products for the base case are shown below. HCl, SO_3, and ash are collected in the chloride scrubber and disposed of in the ash pond.

Component	Kilograms per Hour	Pounds per Hour
Product acid		
98% H_2SO_4	16,300	35,900
Chloride purge		
HCl	290	630
SO_3	40	80
Ash	150	330

The process is evaluated on the basis of 30-day storage of 98% sulfuric acid by-product. Two 3,250 m^3 (858,000 gal) carbon steel tanks have been provided for the base-case design.

Sodium Sulfite (Wellman-Lord) Process

Process Description: The Wellman-Lord process is a sodium sulfite-based scrubbing process that produces SO_2. The SO_2 can be liquefied, or processed to sulfuric acid or elemental sulfur. The system evaluated is Wellman-Lord scrubbing combined with a sulfuric acid plant. An additional case variation is included that uses an Allied Chemical methane reduction unit for sulfur production. The flue gas from the common plenum enters a chloride scrubber where the gas is cooled and the chlorides are removed. From this scrubber the flue gas passes countercurrently to a recirculating sodium sulfite solution in a three-stage valve tray absorber where the SO_2 reacts with the sodium sulfite to form sodium bisulfite and a very small amount of sodium sulfate. Liquor is individually recirculated in each of the three stages to maintain efficient mass transfer. Each absorber has a chevron-type entrainment separator to control entrainment carryover in the gas stream.

A portion of the scrubber effluent is processed to remove Na_2SO_4 by evaporation and selective crystallization in a steam-heated, forced-circulation evaporator serving all four scrubber trains. The clear overflow, enriched in $NaHSO_3$, is returned to the regeneration area. The bottoms, consisting of a slurry enriched in Na_2SO_4 crystals, are centrifuged to produce a solid containing about two-thirds Na_2SO_4 and one-third Na_2SO_3. The centrate is returned to the regeneration area; the solids are dried in a steam-heated dryer and conveyed to a storage silo for sale or discard. There is a potential market available in the paper industry for this material.

The regeneration system consists of two trains of double-effect, forced-circulation evaporators. Scrubber effluent, combined with liquid from the sulfate removal process, is heated and 60% is pumped to the first-effect evaporators and 40% to the second-effect evaporators. The first effect is steam heated; the second effect is heated by combined first-effect vapor and sulfate crystallizer vapor. Some $Na_2S_2O_3$ formed in the first-effect evaporator is removed by a purge stream. Evaporator and stripper overhead vapor, containing H_2O and SO_2, is dried and compressed. The concentrated SO_2 stream is routed to a processing plant. SO_2-bearing condensate from the second-effect evaporator heater, the condensers, and the compressor is steam-stripped and combined with evaporator bottoms. The resulting mixture, enriched in Na_2SO_3, is returned to the absorber system.

The concentrated SO_2 stream is converted to H_2SO_4 in a single-contact, single-absorption acid plant. The tail gas containing unreacted SO_2 is returned to the scrubber. The Allied Chemical methane reduction process is used for a sulfur production case variation. In this process, about 60% of the SO_2 from the Wellman-Lord plant is reduced directly to elemental sulfur by reaction with methane in a primary reduction reactor. Gas leaving the primary reactor, containing H_2S and SO_2 in a molar ratio of 2:1, is routed to a catalytic Claus converter where total conversion of SO_2 to S is increased to about 95%. Sulfur, condensed from the gas stream after the primary and secondary reactors, is pumped to storage. Noncondensed gases from the last sulfur condenser are pumped to an incinerator to oxidize remaining sulfur compounds back to SO_2. After incineration the gases are recycled back to the SO_2 absorbers.

Specific Process Premises:

- The flue gas is cooled from 149°C (300°F) to 54°C (130°F) in the chloride scrubber at an L/G ratio of 1.3 ℓ/m^3 (10 gal/10^3 aft^3).
- A three-stage valve tray absorber with two chimney trays and a superficial velocity of 3 m/sec (10 ft/sec), an L/G ratio of 0.4 ℓ/m^3 (3 gal/10^3 aft^3), and a pressure drop, including the mist eliminator of 2.9 kPa (11.6 inches H_2O).
- Stoichiometry is 2.0 mols of Na_2CO_3 to 1.0 mol of Na removed in the sulfate purge stream.

Energy Requirements: For base-case conditions, reheat of the cleaned gas requires 36.2 x 10^3 kg/hr (79,700 lb/hr) of 243°C (470°F) steam at 3.55 x 10^3 kPa absolute pressure (500 psig), equivalent to about 15.1 x 10^6 kcal/hr. In the purge area, the sulfate dryer uses 0.7 x 10^3 kg/hr (1,600 lb/hr) of 243°C (470°F) saturated steam and the sulfate crystallizer uses 10.3 x 10^3 kg/hr (22,800 lb/hr) of 121°C (250°F) steam at 2.10 x 10^3 kPa (15 psig) absolute pressure. Total consumption in the purge area is equivalent to 6.1 x 10^6 kcal/hr. Steam consumption for the first-effect evaporators and the stripper in the regeneration area is

70.4 kg/hr (155,100 lb/hr) of 121°C (250°F) saturated steam, equivalent to 40.5×10^6 kcal/hr. The acid plant produces 3.6×10^3 kg/hr (7,900 lb/hr) of 121°C (250°F) saturated steam, equivalent to 1.98×10^6 kcal/hr. This is taken into account as a heat credit for the process.

The electrical power demand for the base case is about 11,300 kW or 2.3% of the rated capacity of a 500-MW power plant. For 6,000 hours of operation, the annual electrical energy consumption is 68.1×10^6 kWh.

The total equivalent energy consumption for the base case is approximately 92.35×10^6 kcal/hr or 8.1% of the input energy required for the 500-MW power unit.

By-Product Management: Electrostatic precipitators remove 99.2% of the fly ash in the flue gas. Most of the remainder is removed in the chloride scrubber; therefore, it is assumed that the by-products contain little, if any, fly ash. (Fly ash emission from oil-fired units does not exceed the EPA particulate emission standard and fly ash collection facilities are not included in oil-fired power plant design.) Projected mass flow rates of by-product streams for the base case are shown below.

Component	Kilograms per Hour	Pounds per Hour
Sulfuric acid		
98% H_2SO_4	15,100	33,300
Sodium sulfate		
Na_2SO_4	680	1,500
Na_2SO_3	450	1,000
Other Na salts	50	100
Total	1,180	2,600
Chloride purge		
HCl	290	630
SO_3	40	80
Ash	150	330

The by-product sulfuric acid is stored in a tank of 30-day capacity until sold. For the purposes of this study the sodium sulfate is assumed to be sold.

Carbon Adsorption Process

Process Description: The Bergbau-Forschung/Foster Wheeler process uses dry carbon adsorption followed by thermal regeneration to reactivate the adsorbent. A four-train adsorption system, each with a two-stage adsorber, is used. The adsorber stages are separate but adjacent vessels, each containing louvered moving beds of activated char.

Water is mixed with the flue gas to obtain an adsorber inlet temperature of 121°C (250°F). The gas then moves in horizontal crossflow through a vertical char bed where SO_2, SO_3, H_2O, and O_2 are adsorbed. Although some NO_X removal has been reported, the process is not designed for NO_X removal. The char bed also filters out some residual fly ash in the flue gas, which, leaving the first stage, is divided into upper and lower flow streams. The upper flows to the first-stage adsorber ID fan and then to the stack plenum, the lower, to the second-stage adsorber and a separate ID fan. Because of exothermic reactions the gases are exhausted about 14°C above inlet temperature, eliminating need for reheat.

The char moves by gravity in the adsorber beds. The velocity ranges from 0.3 to 0.91 m/hr (1 to 3 ft/hr). Char leaving the adsorbers is saturated with H_2SO_4 formed by reaction between SO_2, O_2 and H_2O. The saturated char is screened to remove fly ash and char fines and transported to the regeneration area. Saturated char and hot sand are gravity-fed into the regenerator vessel which operates with a reducing atmosphere. Sand enters the regenerator at 815°C (1500°F) and char enters at the flue gas temperature. The saturated char is heated to 650°C (1200°F) to liberate the adsorbed gases and return the char to its activated state. Some char reacts with oxygen in the gas and is chemically consumed. Sand and regenerated char leave the regenerator at 650°C (1200°F) and are separated by screens. The char is then routed to the char cooling area and the sand is heated and returned to the regenerator.

In the char cooling area, the char is cooled in two stages. First-stage cooling by indirect heat exchange with cleaned flue gas lowers char temperature to 200°C. The temperature is reduced to 120°C in the second stage by direct-contact water spray. The steam produced is unsuitable for use by the steam plant. It can, however, be used to preheat boiler feedwater. Makeup char is added to replace losses. The char is then transported back to the top of the adsorbers. The sand is simultaneously heated and pneumatically elevated to the regenerator by combustion gases. The heater gas may be routed either to the stack or to the adsorbers, depending on whether or not it contains SO_2.

SO_2 reduction is accomplished by the Foster Wheeler Resox process. The SO_2-rich gas is reacted with rice-sized anthracite coal to produce gaseous elemental sulfur. SO_2 and air are injected at the bottom of a Resox reactor containing downward-moving coal. The operating temperature is 800° to 850°C in the main reaction zone. In addition to sulfur vapor, some H_2S, COS, and CS_2 are formed in the reactor. The product gases, flowing out the top of the reactor at 350° to 400°C, are routed to a shell and tube exchanger to condense the sulfur product and produce steam. Recovered sulfur is then collected and pumped to storage. Noncondensed gases from the sulfur condenser are pumped to an incinerator to oxidize the remaining sulfur compounds back to SO_2. After incineration, the gases are cooled in a steam-producing waste heat boiler and recycled back to the SO_2 adsorbers.

Specific Process Premises:

- The adsorbers have superficial velocities of 7.6 m/sec (25 ft/sec) in channels between the char beds and 0.3 m/sec (1 ft/sec) approaching the char beds. Pressure drop is 0.5 kPa (2.0 inches H_2O) in the first stage and 0.3 kPa (1 inch H_2O) in the second stage.
- The saturated char SO_2 loading is 7.5 kg of sulfur to 100 kg of char.

Energy Requirements: For base-case conditions, steam consumption in the SO_2 reduction and sulfur storage area totals 1,300 kg/hr (2,860 lb/hr) of 148°C (298°F) steam at 450 kPa absolute pressure (50.3 psig) equivalent to about 0.66×10^6 kcal/hr. In addition to the steam consumed in this area, steam is produced in some areas. This, along with heat used to preheat boiler feedwater, is taken into account as a heat credit. The waste heat boiler produces 4,600 kg/hr (10,100 lb/hr) of 121°C (250°F) steam at 210 kPa absolute pressure (15.1 psig) equivalent to about 2.79×10^6 kcal/hr. The sulfur production unit produces 17,400 kg/hr (38,300 lb/hr) of 121°C (250°F) steam at 210 kPa absolute pressure (15.1 psig) equivalent to about 10.58×10^6 kcal/hr. The waste heat from the char cooler is used to preheat boiler feedwater from 38°C (100°F) to 88°C (190°F). This is equivalent to about 1.45×10^6 kcal/hr.

The electrical power demand for the base-case carbon adsorption process is about 2,650 kW or 0.53% of the rated capacity of a 500-MW power plant. For 6,000 hours of operation the annual electrical energy consumption is 15.92 x 10^6 kWh.

Fuel oil is used as a source of energy at an annual rate of 15.48 x 10^6 liters (4.09 x 10^6 gallons), equivalent to about 22.54 x 10^6 kcal/hr, to heat sand. For SO_2 reduction the annual consumption is 3.12 x 10^6 liters (0.82 x 10^6 gallons). The total fuel oil consumption for energy is equivalent to 27.10 x 10^6 kcal/hr. There is no reheat steam required for this process.

The total equivalent energy consumption for the base case is approximately 19.02 x 10^6 kcal/hr or 1.68% of the input energy required for the 500-MW power unit.

By-Product Management: ESP units remove 99.2% of the fly ash from the flue gas; some additional fly ash is removed with the char fines. The char fines and Resox waste may be burned in the boiler. These two materials have been assigned a monetary value based on their heating values and are treated as saleable by-products in the determination of net annual revenue requirements. (Fly ash emission from oil-fired units does not exceed the EPA particulate emission standard and fly ash collection facilities are not included in oil-fired power plant design.) Projected mass flow rates of by-product streams are shown below.

Component	Kilograms per Hour	Pounds per Hour
Sulfur	4,900	10,900
Char fines	1,660	3,660
Fly ash	150	330
Resox waste	2,600	5,700
Total	4,410	9,690

RESULTS

Process and economic evaluation results for the nine FGD processes evaluated are summarized in Tables 12.1 through 12.10. Figures 12.1 and 12.2 show the economic results geographically, illustrating the accuracy ranges projected for each process. Capital costs are in mid-1979 U.S. dollars. Revenue requirements are in mid-1980 U.S. dollars.

Because the evaluations are based on generalized premises and limited actual cost data are available for some of the processes, no attempt is made to compare the projected costs with costs for actual site-specific installations.

Table 12.1: Limestone Sludge Process Material, Energy, and Cost Summary

Fuel	Coal	Coal	Coal	Coal	Lignite	Oil
Percent Sulfur	0.8	1.4	2.0	3.5	0.5	2.5
Raw materials, kg/hr						
Limestone	6,000	8,900	14,000	25,500	4,300	12,000
End product						
Sludge						
m³/hr	13.3	19.77	31.35	56.93	9.70	26.73
ft³/hr	470.0	698.0	1,107.1	2,010.4	342.4	943.9
kg/hr	17,561	26,079	41,364	75,112	12,792	35,265
lb/hr	38,715	57,493	91,190	165,590	28,200	77,744
wt % solids	40	40	40	40	40	40
Energy						
Reheat, kcal/hr	18.06 M	17.45 M	17.53 M	17.62 M	18.52 M	14.67 M
Electricity, kcal/hr	16.98 M	16.66 M	17.14 M	18.13 M	17.27 M	14.40 M
Total						
kcal/hr	35.04 M	34.11 M	34.67 M	35.75 M	35.79 M	29.07 M
Btu/hr	139.2 M	135.6 M	137.5 M	141.9 M	141.9 M	115.5 M
% input	3.27	3.18	3.23	3.33	3.34	2.71
Land						
Process, ha	3.7	3.7	3.7	3.7	3.7	3.7
Waste disposal, ha	38.6	51.8	74.6	123.0	29.8	65.8
Total						
ha	42.3	55.5	78.3	126.7	33.5	69.5
acres	104.5	137.1	193.5	313.1	82.8	171.7
Costs*, $						
Capital investment	39,995,000	41,634,000	45,516,000	53,083,000	39,036,000	40,643,000
Low range	34,112,000	35,526,000	38,863,000	45,376,000	33,285,000	34,705,000
High range	47,838,000	49,778,000	54,387,000	63,358,000	46,704,000	48,559,000
Annual revenue requirements	11,151,700	11,538,200	12,574,200	14,638,000	10,945,400	11,385,400
Low range	9,933,800	10,279,000	11,206,400	13,064,500	9,752,100	10,168,000
High range	12,775,200	13,219,500	14,397,200	16,736,100	12,536,500	13,008,300

*Accuracy range: +20 to -15%.

Source: EPA-600/7-80-142

Table 12.2: Limestone Sludge Process Size Variations, Material, Energy, and Cost Summary

MW	200	500	700	1,000
Fuel	Coal	Coal	Coal	Coal
Percent sulfur	3.5	3.5	3.5	3.5
Raw materials, kg/hr				
Limestone	10,400	25,500	35,300	49,300
End product				
Sludge				
m^3/hr	23.28	56.93	78.82	110.07
ft^3/hr	822.0	2,010.4	2,783.1	3,886.7
kg/hr	30,713	75,112	103,983	145,216
lb/hr	67,710	165,590	229,240	320,140
wt % solids	40	40	40	40
Energy				
Reheat, kcal/hr	7.21 M	17.62 M	24.39 M	34.07 M
Electricity, kcal/hr	7.72 M	18.13 M	24.74 M	33.68 M
Total				
kcal/hr	14.93 M	35.75 M	49.13 M	67.75 M
Btu/hr	59.2 M	141.79 M	194.95 M	268.83 M
% input	3.39	3.33	3.30	3.26
Land				
Process, ha	3.1	3.7	3.8	4.2
Waste disposal, ha	61.0	123.0	159.0	209.0
Total				
ha	64.1	126.7	162.8	213.2
acres	158.3	313.1	402.3	526.9
Costs*, $				
Capital investment	26,302,000	53,083,000	70,967,000	93,941,000
Low range	22,481,000	45,376,000	60,657,000	80,296,000
High range	31,397,000	63,358,000	84,713,000	112,135,000
Annual revenue requirements	7,029,100	14,638,000	19,734,300	26,384,700
Low range	6,260,400	13,064,500	17,619,200	23,573,800
High range	8,054,000	16,736,100	22,554,300	30,132,600

*Accuracy range: +20 to -15%.

Source: EPA-600/7-80-142

Table 12.3: Lime Sludge Process Material, Energy, and Cost Summary

Fuel Percent Sulfur	Coal 0.8	Coal 1.4	Coal 2.0	Coal 3.5	Lignite 0.5	Oil 2.5
Raw materials, kg/hr						
Lime	2,400	3,500	5,600	10,200	1,700	4,800
End product						
Sludge						
m^3/hr	10.22	15.17	24.07	43.70	7.44	20.52
ft^3/hr	360.3	535.0	848.6	154.1	262.4	723.4
kg/hr	13,394	19,891	31,549	57,290	9,756	26,898
lb/hr	29,529	43,851	69,553	126,300	21,509	59,298
wt % solids	40	40	40	40	40	40
Energy						
Reheat, kcal/hr	18.00 M	17.39 M	17.47 M	17.56 M	18.46 M	14.62 M
Electricity, kcal/hr	16.82 M	16.34 M	16.54 M	16.89 M	17.21 M	13.88 M
Total						
kcal/hr	34.82 M	33.73 M	34.01 M	34.45 M	35.67 M	28.50 M
Btu/hr	138.2 M	133.9 M	135.0 M	136.7 M	141.5 M	113.1 M
% input	3.25	3.14	3.17	3.21	3.33	2.51
Land						
Process, ha	2.5	2.5	2.5	2.5	2.5	2.5
Waste disposal, ha	32.5	44.8	63.5	104.0	25.5	54.9
Total						
ha	35.0	47.3	66.0	106.5	28.0	57.4
acres	86.5	116.9	163.1	262.4	69.2	141.8
Costs*, $						
Capital investment	37,795,000	38,912,000	41,952,000	47,743,000	37,165,000	37,286,000
Low range	32,244,000	33,216,000	35,841,000	40,853,000	31,695,000	31,855,000
High range	45,196,000	46,507,000	50,099,000	56,932,000	44,458,000	44,526,000
Annual revenue requirements	11,064,200	11,498,200	12,644,700	14,972,100	10,836,300	11,387,700
Low range	9,912,000	10,320,100	11,386,000	13,562,700	9,698,200	10,271,300
High range	12,600,400	13,069,200	14,323,000	16,851,500	12,353,800	12,876,100

*Accuracy range: +20 to -15%.

Source: EPA-600/7-80-142

Table 12.4: Double-Alkali Sludge Process Material, Energy, and Cost Summary

Fuel Percent Sulfur	Coal 0.8	Coal 1.4	Coal 2.0	Coal 3.5	Lignite 0.5	Oil 2.5
Raw materials, kg/hr						
Lime	2,277	3,381	5,363	9,737	1,659	4,572
Soda ash	186	277	439	797	136	374
End product						
Sludge						
m^3/hr	6.98	10.37	16.45	29.87	5.09	14.03
ft^3/hr	246.6	366.2	580.9	1,054.9	179.6	495.3
kg/hr	9,075	13,476	21,375	38,815	6,610	18,223
lb/hr	20,006	29,710	47,123	85,570	14,573	40,175
wt % solids	55	55	55	55	55	55
Energy						
Reheat, kcal/hr	18.06 M	17.45 M	17.53 M	17.62 M	18.52 M	14.67 M
Electricity, kcal/hr	8.08 M	7.99 M	8.33 M	9.03 M	8.18 M	7.02 M
Total						
kcal/hr	26.14 M	25.44 M	25.86 M	26.65 M	26.70 M	21.69 M
Btu/hr	103.7 M	101 M	102.6 M	105.8 M	106 M	86.1 M
% input	2.48	2.41	2.45	2.52	2.54	1.91
Land						
Process, ha	4.9	4.9	4.9	4.9	4.9	4.9
Waste disposal, ha	33.0	44.0	62.0	99.0	31.1	48.0
Total						
ha	37.9	48.9	66.9	103.9	36.0	52.9
acres	150.4	194.0	265.5	412.3	89.1	209.9
Costs*, $						
Capital investment	40,537,000	42,179,000	46,045,000	53,231,000	39,595,000	40,659,000
Low range	32,581,000	33,926,000	37,075,000	43,004,000	31,817,000	32,733,000
High range	52,472,000	54,559,000	59,500,000	68,662,000	51,261,000	52,546,000
Annual revenue requirements	11,173,000	11,793,600	13,080,000	16,010,600	10,828,000	11,879,500
Low range	9,528,500	10,091,700	11,258,600	14,359,800	9,217,700	10,251,200
High range	13,639,700	14,346,600	15,811,900	19,169,400	13,234,300	14,321,700

*Accuracy range: +30 to -20%.

Source: EPA-600/7-80-142

Table 12.5: Seawater Process Material, Energy, and Cost Summary

Fuel Percent sulfur	Coal 1.4	Coal 0.8	Lignite 0.5	Oil 2.5	Low Reheat Coal 1.4
End product					
Seawater					
ℓ/hr	71,385,400	51,526,200	35,692,700	83,214,800	71,385,400
kg/hr	72,124,300	52,041,400	36,062,200	84,071,900	72,124,300
lb/hr	158,864,200	114,731,700	79,432,100	185,346,900	158,864,200
Energy					
Reheat, kcal/hr	28.70 M	29.69 M	30.46 M	24.13 M	14.46 M
Electricity, kcal/hr	15.91 M	14.89 M	13.88 M	15.47 M	15.91 M
Total					
kcal/hr	44.61 M	44.58 M	44.34 M	39.60 M	30.37 M
Btu/hr	177.0 M	176.9 M	175.9 M	157.1 M	120.5 M
% input	4.22	4.22	4.21	3.49	2.82
Land					
Process					
ha	2.4	2.4	2.4	2.4	2.4
acres	6	6	6	5	6
Costs*, $					
Capital investment	30,048,000	29,590,000	29,068,000	27,937,000	28,582,000
Low range	21,157,000	20,838,000	20,471,000	19,691,000	20,106,000
High range	44,868,000	44,177,000	43,397,000	41,682,000	42,707,000
Annual revenue requirements	8,707,100	8,667,100	8,516,800	8,575,600	7,837,700
Low range	6,972,700	6,960,500	6,840,700	6,968,800	6,185,700
High range	11,597,700	11,511,800	11,310,300	11,253,400	10,590,900

*Accuracy range: +50 to -30%.

Source: EPA-600/7-80-142

Table 12.6: Lime Gypsum (Saarberg-Hölter) Process Material, Energy, and Cost Summary

Fuel	Coal	Coal	Coal	Coal	Lignite	Oil
Percent Sulfur	0.8	1.4	2.0	3.5	0.5	2.5
Raw materials, kg/hr						
Lime	2,255	3,344	5,297	9,625	1,634	4,525
Formic acid	1.8	2.7	4.3	7.9	1.3	3.7
Flocculant	0.18	0.26	0.42	0.76	0.13	0.36
Nalco	0.02	0.03	0.04	0.08	0.01	0.04
End product						
Gypsum						
m^3/hr	6.06	9.00	14.27	25.92	4.41	12.17
ft^3/hr	214.1	317.6	503.9	915.3	155.7	429.8
kg/hr	8,263	12,260	19,445	35,320	6,008	16,586
lb/hr	18,200	27,000	42,830	77,800	13,230	36,530
wt % solids	80	80	80	80	80	80
Energy						
Reheat, kcal/hr	18.14 M	17.53 M	17.61 M	17.70 M	18.61 M	14.74 M
Electricity, kcal/hr	12.59 M	13.76 M	16.34 M	22.00 M	12.13 M	13.78 M
Total						
kcal/hr	30.73 M	31.29 M	33.95 M	39.70 M	30.74 M	28.52 M
Btu/hr	121.94 M	124.16 M	134.71 M	157.53 M	121.98 M	113.17 M
% input	2.89	2.93	3.17	3.67	2.89	2.51
Land						
Process, ha	4.46	4.46	4.46	4.46	4.46	4.46
By-product storage, ha	0.4	0.4	0.4	0.4	0.4	0.4
Total						
ha	4.86	4.86	4.86	4.86	4.86	4.86
acres	12	12	12	12	12	12
Costs*, $						
Capital investment	35,481,000	36,472,000	39,173,000	44,024,000	34,905,000	34,802,000
Low range	24,988,000	25,701,000	27,631,000	31,114,000	24,574,000	24,557,000
High range	52,970,000	54,424,000	58,407,000	65,541,000	52,124,000	51,878,000
Annual revenue requirements	9,825,900	10,303,700	11,443,300	13,706,300	9,571,100	10,347,600
Low range	7,791,500	8,214,400	9,201,300	11,192,900	7,568,900	8,362,400
High range	13,216,300	13,785,800	15,179,800	17,895,300	12,908,000	13,656,200

*Accuracy range: +50 to -30%.

Source: EPA-600/7-80-142

Table 12.7: Jet-Bubbling Limestone (Chiyoda Thoroughbred 121) Process Material, Energy, and Cost Summary

Fuel Percent Sulfur	Coal 0.8	Coal 1.4	Coal 2.0	Coal 3.5	Lignite 0.5	Oil 2.5
Raw materials, kg/hr						
Limestone	4,067	6,048	9,586	17,400	2,964	8,180
End product						
Gypsum						
m^3/hr	6.19	9.18	14.57	26.45	4.50	12.42
ft^3/hr	218.42	324.30	514.42	934.14	159.06	438.54
kg/hr	8,421	12,503	19,834	36,015	6,132	16,908
lb/hr	18,565	27,565	43,725	79,400	13,520	37,275
wt % solids	80	80	80	80	80	80
Energy						
Reheat, kcal/hr	18.61 M	17.98 M	18.07 M	18.16 M	19.09 M	15.12 M
Electricity, kcal/hr	17.32 M	16.97 M	17.42 M	18.35 M	16.63 M	14.64 M
Total						
kcal/hr	35.93 M	34.95 M	35.49 M	36.51 M	35.72 M	29.76 M
Btu/hr	142.57 M	138.68 M	140.82 M	144.87 M	141.74 M	118.09 M
% input	3.35	3.26	3.31	3.40	3.34	2.62
Land						
Process, ha	4.46	4.46	4.46	4.46	4.46	4.46
By-product storage, ha	0.4	0.4	0.4	0.4	0.4	0.4
Total						
ha	4.86	4.86	4.86	4.86	4.86	4.86
acres	12	12	12	12	12	12
Costs*, $						
Capital investment	42,002,000	42,007,000	43,819,000	47,017,000	42,095,000	38,532,000
Low range	29,538,000	29,545,000	30,826,000	33,091,000	29,602,000	27,118,000
High range	62,774,000	62,776,000	65,474,000	70,227,000	62,917,000	57,556,000
Annual revenue requirements	10,948,500	10,919,500	11,360,700	12,160,100	10,998,700	10,219,200
Low range	8,524,700	8,496,000	8,831,900	9,446,000	8,568,900	8,003,400
High range	14,988,200	14,958,600	15,575,600	16,683,500	15,048,200	13,912,300

*Accuracy range: +50 to -30%

Source: EPA-600/7-80-142

Table 12.8: Magnesium Oxide Process Material, Energy, and Cost Summary

Fuel Percent Sulfur	Coal 0.8	Coal 1.4	Coal 2.0	Coal 3.5	Lignite 0.5	Oil 2.5
Raw materials						
MgO, kg/hr	51	76	119	218	38	103
Catalyst, ℓ/hr	0.1	0.1	0.2	0.3	0.1	0.1
Agricultural limestone, kg/hr	431	832	1,255	420	398	0
End product						
Sulfuric acid						
ℓ/hr	2,041	3,026	4,800	8,717	1,485	4,090
kg/hr	3,735	5,538	8,785	15,951	2,717	7,484
lb/hr	8,233	12,210	19,367	35,167	5,990	16,500
*wt.% H_2SO_4	100	100	100	100	100	100
Energy						
Reheat, kcal/hr	18.57 M	17.95 M	18.03 M	18.12 M	19.05 M	15.09 M
Electricity, kcal/hr	16.38 M	16.57 M	17.87 M	20.66 M	16.38 M	12.94 M
Oil, kcal/hr	7.90 M	11.73 M	18.60 M	33.78 M	5.75 M	15.86 M
Heat credit, kcal/hr	(0.84 M)	(1.25 M)	(1.98 M)	(3.60 M)	(0.61 M)	(1.69 M)
Total						
kcal/hr	42.01 M	45.00 M	52.52 M	68.96 M	40.57 M	42.20 M
Btu/hr	166.7 M	178.6 M	208.4 M	273.6 M	161.0 M	167.4 M
% input	3.89	4.14	4.81	6.26	3.58	3.72
Land						
Process						
ha	4.86	4.86	4.86	4.86	4.86	4.86
acres	12	12	12	12	12	12
Costs*, $						
Capital investment	48,926,000	52,043,000	58,108,000	68,434,000	47,115,000	44,591,000
Low range	39,252,000	41,757,000	46,632,000	54,934,000	37,798,000	35,799,000
High range	63,436,000	67,472,000	75,324,000	88,685,000	61,090,000	57,779,000
Annual revenue requirements	12,949,500	13,651,200	15,114,300	17,546,900	12,573,900	11,984,600
Low range	10,999,500	11,576,400	12,759,400	14,809,500	10,697,100	10,208,300
High range	15,874,400	16,763,000	18,592,700	21,652,700	15,389,100	14,646,600

*Accuracy range: +30 to -20%.

Source: EPA-600/7-80-142

Table 12.9: Sodium Sulfite (Wellman-Lord) Process Material, Energy, and Cost Summary

Fuel Percent sulfur	Coal 0.8	Coal 1.4	Coal 2.0	Coal 3.5	Lignite 0.5	Oil 2.5	Sulfur Production Coal 3.5
Raw materials							
Sodium carbonate, kg/hr	218	323	512	930	159	437	930
Catalyst, ℓ/hr	0.07	0.10	0.16	0.29	0.05	0.14	–
Agricultural limestone, kg/hr	431	832	1,254	420	398	–	420
Filter aid, kg/hr	2	4	6	11	2	5	11
Natural gas, m^3/hr	–	–	–	–	–	–	2,311
End product							
100% sulfuric acid							
ℓ/hr	[illegible]	[illegible]	4,446	8,073	1,375	3,790	–
kg/hr	[illegible]	[illegible]	8,136	14,774	2,516	6,935	–
lb/hr	[illegible]	[illegible]	17,937	32,570	5,546	15,290	–
Sodium sulfate							
kg/hr	[illegible]	410	650	1,181	201	555	1,181
lb/hr	610	903	1,433	2,603	443	1,223	2,603
Sulfur							
kg/hr	–	–	–	–	–	–	4,802
lb/hr	–	–	–	–	–	–	10,857
Energy							
Reheat, kcal/hr	15.47 M	14.95 M	15.02 M	15.10 M	15.87 M	12.57 M	15.10 M
Process steam, kcal/hr	10.90 M	16.19 M	25.68 M	46.64 M	7.94 M	21.90 M	47.20 M
Electricity, kcal/hr	17.99 M	18.68 M	20.90 M	25.73 M	17.72 M	15.44 M	25.59 M
Natural gas, kcal/hr	–	–	–	–	–	–	20.56 M
Heat credit, kcal/hr	(0.46 M)	(0.69 M)	(1.09 M)	(1.98 M)	(0.34 M)	(0.93 M)	(2.79 M)
Total - kcal/hr	43.90 M	49.13 M	60.51 M	85.49 M	41.19 M	48.98 M	105.66 M
Btu/hr	174.2 M	194.9 M	240.1 M	339.2 M	163.4 M	194.4 M	419.3 M
% input	4.13	4.64	5.73	8.14	3.78	4.53	9.93
Land							
Process - ha	3.2	3.2	3.2	3.2	3.2	3.2	3.2
acres	8	8	8	8	8	8	8
Costs*, $							
Capital investment	46,836,000	50,307,000	56,939,000	68,722,000	44,837,000	44,215,000	71,342,000
Low range	37,581,000	40,372,000	45,706,000	55,187,000	35,975,000	35,504,000	57,343,000
High range	60,719,000	65,210,000	73,789,000	89,025,000	58,131,000	57,280,000	92,341,000
Annual revenue requirements	12,218,300	13,081,000	14,802,200	17,886,400	11,754,700	11,801,900	21,015,700
Low range	10,434,500	11,164,000	12,629,900	15,257,400	10,048,100	10,124,600	18,296,500
High range	14,893,900	15,956,500	18,060,200	21,829,800	14,314,400	14,317,700	24,094,700

*Accuracy range: +30 to -20%.

Source: EPA-600/7-80-142

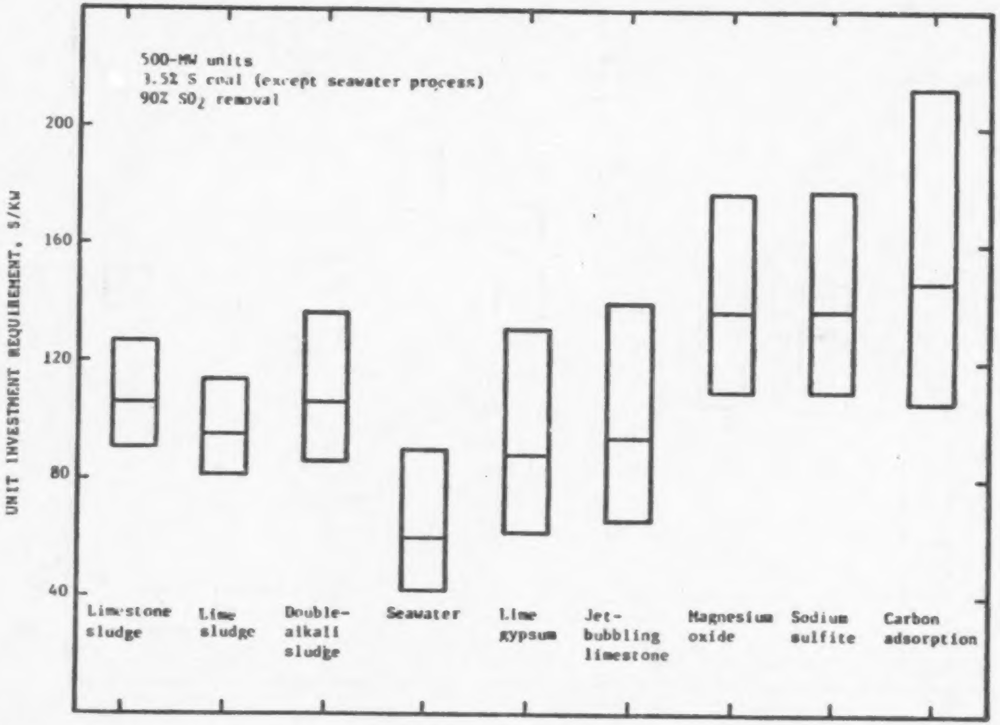
Table 12.10: Carbon Adsorption Process Material, Energy, and Cost Summary

Fuel Percent Sulfur	Coal 0.8	Coal 1.4	Coal 2.0	Coal 3.5	Lignite 0.5	Oil 2.5
Raw materials						
Sand, kg/hr	106	151	257	454	76	212
Char, kg/hr	544	816	1,300	2,359	408	1,104
Anthracite coal, kg/hr	1,179	1,754	1,255	5,035	862	2,359
End product						
Sulfur						
lb/hr	2,533	3,800	6,000	10,900	1,867	5,133
kg/hr	1,149	1,724	2,752	4,944	847	2,328
Char fines and Resox waste						
lb/hr	2,267	3,400	5,366	9,733	1,633	4,567
kg/hr	1,028	1,542	2,434	4,415	741	2,071
Energy						
Electricity, kcal/hr	3.86 M	4.08 M	3.83 M	6.01 M	3.74 M	3.99 M
Process steam, kcal/hr	0.15 M	0.23 M	0.36 M	0.66 M	0.11 M	0.31 M
Fuel oil, kcal/hr	6.33 M	9.41 M	14.92 M	27.10 M	4.61 M	12.72 M
Heat credit, kcal/hr	(3.46 M)	(5.14 M)	(8.16 M)	(14.82 M)	(2.52 M)	(6.96 M)
Total						
kcal/hr	6.88 M	8.58 M	10.95 M	18.95 M	5.94 M	10.06 M
Btu/hr	27.3 M	34.0 M	43.4 M	75.2 M	23.6 M	39.9 M
% input	0.61	0.76	0.97	1.68	0.52	0.89
Land						
Process						
ha	4.86	4.86	4.86	4.86	4.86	4.86
acres	12	12	12	12	12	12
Costs*, $						
Capital investment	51,195,000	54,220,000	60,834,000	73,511,000	49,485,000	53,730,000
Low range	33,109,000	39,533,000	44,350,000	53,617,000	36,153,000	39,104,000
High range	69,978,000	78,698,000	88,308,000	106,666,000	71,704,000	78,105,000
Annual revenue requirements	13,899,100	15,803,600	19,982,700	28,489,400	12,780,100	17,542,000
Low range	10,747,500	12,844,600	16,658,600	24,473,000	10,094,500	14,600,600
High range	18,174,700	20,735,200	25,522,900	35,183,500	17,256,100	22,445,100

*Accuracy range: +50 to -30%.

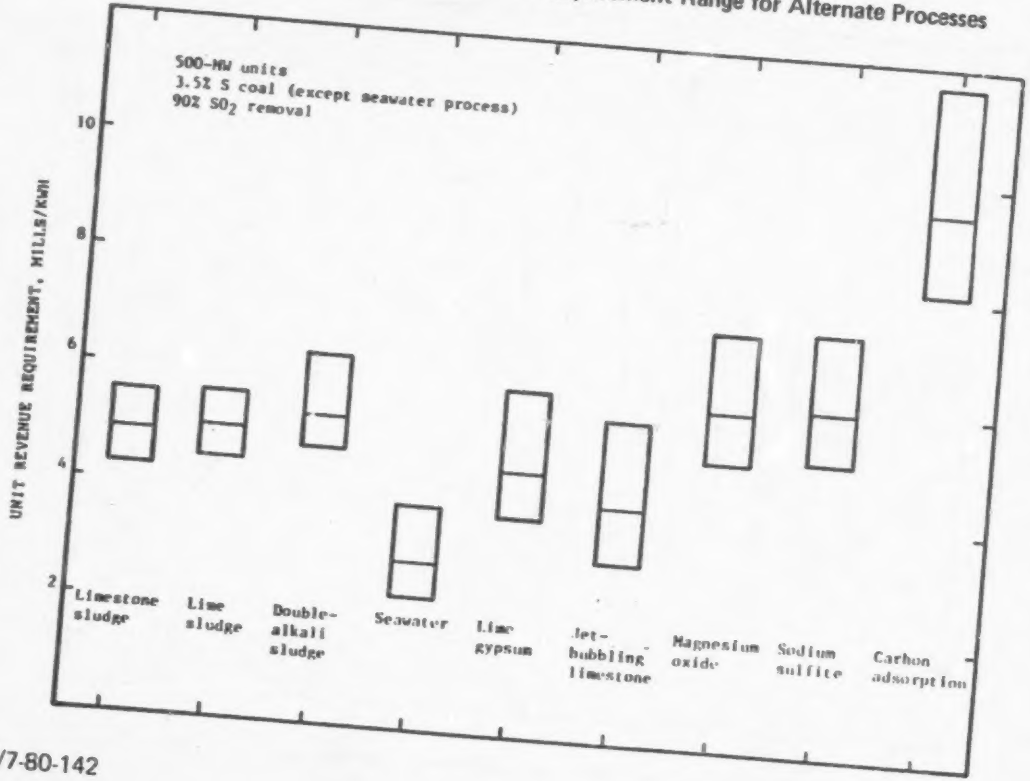
Source: EPA-600/7-80-142

Figure 12.1: Base-Case Unit Investment Range for Alternate Processes



Source: EPA-600/7-80-142

Figure 12.2: Base-Case Unit Revenue Requirement Range for Alternate Processes



Source: EPA-600/7-80-142

This Environmental Protection Agency (EPA) summary lists all of the operational and planned FGD systems in the U.S., as of September 1981, and describes the process type, the supplier, and the sulfur content of fuel being burned. Of special interest is Table 5, (p. 348) which gives the capital and annual operating costs of several FGD processes. Comparative costs are also given for new and retrofit facilities.

United States
Environmental Protection
Agency

Industrial Environmental Research
Laboratory
Research Triangle Park NC 27711

Research and Development

EPA-600/S7-81-012e August 1982

EPA

Project Summary

EPA Utility FGD Survey July—September 1981

M. P. Smith, M. Melia, and N. Gregory

The report, generated by a computerized data base system, presents a survey of operational and planned domestic utility flue gas desulfurization (FGD) systems. It summarizes information contributed by the utility industry, system and equipment suppliers, system designers, research organizations, and regulatory agencies. It presents data on system design, fuel characteristics, operating history, and actual performance. Unit by unit dependability parameters are included and problems and solutions associated with the boilers, scrubbers, and FGD systems are discussed.

The FGD systems are tabulated alphabetically by development status (operational, under construction, or in the planning stages), utility company, system supplier, process, waste disposal practice, and regulatory class. FGD system economic data, definitions, and a glossary of terms are appended to the report. Current data for domestic FGD systems show 92 systems in operation, 42 systems under construction, and 88 planned systems. Projected 1999 FGD controlled capacity in the U.S. is 107,197 MW.

This Project Summary was developed by EPA's Industrial Environmental Research Laboratory, Research Triangle Park, NC, to announce key findings of the research project that is fully documented in a separate report of the same title (see Project Report ordering information at back).

Introduction

The quarterly FGD survey report is generated by a computerized data system known as the Flue Gas Desulfurization Information System (FGDIS). A structure diagram of the FGDIS is illustrated in Figure 1, which presents the informational areas addressed in the system and some representative data items contained in each. The design data contained in the system encompass the entire emission control system and the power generating unit to which it is applied. Performance data for the operational FGD systems include monthly dependability parameters, along with service time and problem/solution descriptions.

In addition to generating the survey report, the FGDIS is available for remote terminal access. Because the report is available only through purchase from NTIS, the data base is the most immediate method for examining the data acquired under the survey program. Access to the FGDIS also enables users to obtain data that are too specific for inclusion in the quarterly report. Information concerning access to the FGDIS can be obtained from Walter Finch, NTIS, 5285 Port Royal Road, Springfield, VA 22161; (703) 487-4808. Custom searches of the FGDIS data can also be arranged.

Executive Summary

Table 1 summarizes the status of FGD systems in the U.S. at the end of September 1981. Table 2 lists the units that have changed status during the third quarter 1981, and Table 3 shows the performance of operating units during this period. The units included in the figures presented in Table 1 are identified in Table 4, and categorical FGD system cost data are presented in Table 5.

Table 1. Number and Total Capacity of FGD Systems

Status	No. of units	Total controlled capacity, MW*	Equivalent scrubbed capacity, MW†
Operational	92	34,937	31,738
Under construction	42	18,226	17,457
Planned:			
Contract awarded	16	9,385	9,169
Letter of intent	11	8,293	8,235
Requesting/evaluating bids	10	5,630	5,630
Considering only FGD systems	51	30,726	30,398
TOTAL	222	107,197	102,627

* The summation of the gross unit capacities (MW) brought into compliance with FGD systems regardless of the percent of the flue gas scrubbed by the FGD system(s).

† The summation of the effective scrubbed flue gas in equivalent MW based on the percent of the flue gas scrubbed by the FGD system(s).

Table 2. Summary of Changes July-September 1981

	Operational		Under construction		Contract awarded		Letter of intent		Requesting/ eval. bids		Considering FGD		Total	
FGD status report	No.	MW[a]	No.	MW[a]	No.	MW[a]	No.	MW[a]	No.	MW[a]	No.	MW[a]	No.	MW[a]
July 31, 1981	88	30,158	40	15,887	21	11,599[b]	11	8,235	19	10,340	48	28,618[b]	227	104,837[b]
Arizona Public Service														
Four Corners 4			+1	755	−1	755								
Four Corners 5			+1	755	−1	755								
Basin Electric Power														
Antelope Valley 2			+1	440	−1	440								
Laramie River 2	+1	570	−1	570										
Central Illinois Light														
Duck Creek 2									−1	450	+1	450		
East Kentucky Power														
J.K. Smith 2							−1	650					−1	650
Kentucky Utilities														
Hancock 1							+1	650	−1	550				
Hancock 2									−1	650	+1	650		
Nevada Power														
Warner Valley 1									−1	295			−1	295
Warner Valley 2									−1	295			−1	295
Northern States Power														
Wisconsin Coal 1											−1	670	−1	670
Pacific Power & Light														
Jim Bridger 1									−1	550	+1	550		
Jim Bridger 2									−1	550	+1	550		
Jim Bridger 3									−1	550	+1	550		
Plains Electric G & T														
Plains Escalante 1			+1	233	−1	233								
San Miguel Electric														
San Miguel 1	+1	400	−1	400										
Seminole Electric														
Seminole 1			+1	620	−1	620								
South Carolina Public Service														
Winyah 4	+1	280	−1	280										

Table 2. (continued)

	Operational		Under construction		Contract awarded		Letter of intent		Requesting/ eval. bids		Considering FGD		Total	
FGD status report	No.	MW[a]	No.	MW[a]	No.	MW[a]	No.	MW[a]	No.	MW[a]	No.	MW[a]	No.	MW[a]
July 31, 1981	88	30,158	40	15,887	21	11,599[b]	11	8,235	19	10,340	48	28,618[b]	227	104,837[b]
Southern Illinois Power														
Marion 5											−1	300	−1	300
Sunflower Electric														
Holcomb 1			+1	347	−1	347								
Utah Power & Light														
Naughton 3	+1	330	−1	330										
West Texas Utilities														
Oklaunion 1					+1	720			−1	720				
Total	92	31,738	42	17,457	16	9,169	11	8,235	10	5,630	51	30,398	222	102,627

[a] Equivalent scrubbed capacity.

[b] This value was modified slightly due to a MW correction.

Table 3. Performance of Operational Units July-September 1981

Plant	FGD system capacity, MW[a]	Flue gas % scrubbed	FGD capacity on line during period MW[a,b]	No information for this period, MW[a]	Shut down throughout period, MW[a]	July 1981 Dependability %[c,e]				August 1981 Dependability %[c,e]				September 1981 Dependability %[c,e]			
						AVL	OPR	REL	UTL	AVL	OPR	REL	UTL	AVL	OPR	REL	UTL
Alabama Electric																	
Tombigbee 2	179	70		179													
Tombigbee 3	179	70		179													
Arizona Electric Power																	
Apache 2	98	50	98			100	91	100	91	100	40	100	36	100	99	100	99
Apache 3	98	50	98			100	69	100	69	100	100	100	100	62	63	62	62
Arizona Public Service																	
Cholla 1	119	100		119													
Cholla 2	264	100		264													
Cholla 4	126	33		126													
Four Corners 1	175	100		175													
Four Corners 2	175	100		175													
Four Corners 3	229	100		229													
Basin Electric Power Coop																	
Laramie River 1	570	100	570			100	95	100	95	97	96	100	92	100	94	100	28
Laramie River 2	570	100	570							96	69	100	68	100	82	100	67
Big Rivers Electric																	
Green 1	242	100	242			98	96	96	96	95	93	93	85	85	96	96	80
Green 2	242	100	242			94	89	89	77	97	94	94	86	100	99	99	98
Central Illinois Light																	
Duck Creek 1	416	100	416			72	63	78	64	78	74	76	74				
Central Illinois Public Service																	
Newton 1	617	100	617			89	70	99	68	100	82	100	82	100	81	100	69
Cincinnati Gas & Electric																	
East Bend 2	650	100	650														
Colorado Ute Electric Association																	
Craig 1	410	90			410	0	0	0	0	0	0	0	0	0	0	0	0
Craig 2	410	90	410			0	0	0	0	33	34	34	33	81	82	83	80
Columbus & Southern Ohio Electric																	
Conesville 5	411	100	411			100	100	100	85	99	83	92	41	94	54	91	8
Conesville 6	411	100	411			99	100	100	74	99	90	93	74	96	86	96	36
Commonwealth Edison																	
Powerton 51	450	100			450				0				0				0

Table 3. (continued)

Plant	FGD system capacity, MW^a	Flue gas % scrubbed	FGD capacity on line during period MW^a,b	No information for this period, MW^a	Shut down throughout period, MW^a	July 1981 Dependability %^c,e				August 1981 Dependability %^c,e				September 1981 Dependability %^c,e			
						AVL	OPR	REL	UTL	AVL	OPR	REL	UTL	AVL	OPR	REL	UTL
Cooperative Power Association																	
Coal Creek 1	327	60	327			37		37			45		36				0
Coal Creek 2	327	60	327			42		37			44		34		60		60
Delmarva Power & Light																	
Delaware City 1	60	100	60			99	99	99	99	77	77	77	47	95	95	95	95
Delaware City 2	60	100	60			77	77	77	77	99	98	98	98	99	96	96	87
Delaware City 3	60	100	60			99	99	99	99	100	100	100	100	90	90	90	90
Duquesne Light																	
Elrama 1-4	510	100	510			99	95	99	95	90	81	98	81				
Phillips 1-6	408	100	408			74	72	73	72	75	64	70	58				
Indianapolis Power & Light																	
Petersburg 3	532	100		532													
Kansas City Power & Light																	
Hawthorn 3	90	100	90			100	100		62	100	100		10	100	100		58
Hawthorn 4	90	100	90			96	100		49	100	100		18	100	100		3
La Cygne 1	820	100	820			97	80	95	64								
Kansas Power & Light																	
Jeffrey 1	540	75	540														
Jeffrey 2	490	70	490														
Lawrence 4	125	100	125														
Lawrence 5	420	100	420														
Kentucky Utilities																	
Green River 1-3	64	100			64	100			0	100			0	100			0
Louisville Gas & Electric																	
Cane Run 4	188	100	188			73	100	100	73	40	66	66	40	61	100	100	61
Cane Run 5	200	100	200			46	100	100	46	50	100	100	50	75	100	100	75
Cane Run 6	299	100	299			67	100	100	67	68	73	73	68	29	100	100	29
Mill Creek 1	358	100	358			75	75	75	72	99	99	99	99	85	96	96	80
Mill Creek 3	427	100	427			63	74	74	63	89	92	92	88	81	81	81	81
Paddy's Run 6	72	100			72	100			0	100			0	100			0
Minnesota Power & Light																	
Clay Boswell 4	475	85	475			100	86	100	75	100	80	100	73	100	89	100	84
Minnkota Power & Light																	
Milton R. Young 2	185	42	185			29	32	33	29	58	49	60	44	100	97	100	86
Monongahela Power																	
Pleasants 1	618	100		618													
Pleasants 2	618	100		618													
Montana Power																	
Colstrip 1	360	100		360													
Colstrip 2	360	100		360													
Montana-Dakota Utilities																	
Coyote 1	440	100	440														
Nevada Power																	
Reid Gardner 1	125	100	125			100	99	100	95	99	99	99	93	96	91	95	77
Reid Gardner 2	125	100	125			99	97	99	78	98	98	98	96	100	99	100	99
Reid Gardner 3	125	100	125			97	96	97	93	98	98	98	98	100	98	100	75
Northern Indiana Public Service																	
Dean H. Mitchell 11	115	99			115	100	0		0	100	0		0	100	0		0
Northern States Power																	
Riverside 6, 7	110	N/A^d		110													
Sherburne 1	740	91	740														
Sherburne 2	740	91	740														
Pacific Power & Light																	
Jim Bridger 4	550	100		550													
Pennsylvania Power																	
Bruce Mansfield 1	917	100	917							97							
Bruce Mansfield 2	917	100	917							97							
Bruce Mansfield 3	917	100	917							99							

Table 3. *(continued)*

Plant	FGD system capacity, MW[a]	Flue gas % scrubbed	FGD capacity on line during period MW[a, b]	No information for this period, MW[a]	Shut down throughout period, MW[a]	July 1981 Dependability %[c, e]				August 1981 Dependability, %[c, e]				September 1981 Dependability %[c, e]			
						AVL	OPR	REL	UTL	AVL	OPR	REL	UTL	AVL	OPR	REL	UTL
Public Service Co. of New Mexico																	
San Juan 1	361	100	361			100	97	100	96	100	100	100	100	100	75	90	32
San Juan 2	350	100	350			93	80	86	80	100	54	100	43	65	19	19	19
San Juan 3	534	100	534			98	96	97	95	86	86	86	85	97	32	32	31
Salt River Project																	
Coronado 1	280	80	280														
Coronado 2	280	80	280														
San Miguel Electric																	
San Miguel 1	400	100	400														
Sikeston Board of Municipal Utilities																	
Sikeston 1	235	100			235				0				0				0
South Carolina Public Service Authority																	
Winyah 2	140	60	140			89	93	96	44	98	98	98	98	96	98	96	88
Winyah 3	280	100	280			88	83	84	82	82	84	82	70	98	96	98	86
Winyah 4	280	100	280														
South Mississippi Electric																	
R.D. Morrow, Sr. 1	124	62	124			100	99	99	94	100	100	100	100	93	99	99	98
R.D. Morrow, Sr. 2	124	62	124			100	97	100	93	100	100	100	96	98	98	98	98
Southern Illinois Power Coop																	
Marion 4	173	100	173			74	74	94	74	52	47	60	47				
Southern Indiana Gas and Electric																	
A.B. Brown 1	265	100	265			91	84	87	84	92	83	89	83	96	97	90	90
Springfield City Utilities																	
Southwest 1	194	100	194			97	93	96	93	98	87	50	87	89	78	86	76
Springfield Water, Light & Power																	
Dallman 3	185	90	185			84	44	44	42	0	0	0	0				
St. Joe Zinc																	
G.F. Weston 1	60	N/A[d]	60			0	0	0	0	0	0	0	0	5	5	5	5
Tennessee Valley Authority																	
Shawnee 10A	10	N/A[d]		10													
Shawnee 10B	10	N/A[d]		10													
Widows Creek 7	575	100	575														
Widows Creek 8	550	100			550				0				0				0
Texas Power & Light																	
Sandow 4	382	70	382														
Texas Utilities																	
Martin Lake 1	595	75	595														
Martin Lake 2	595	75	595														
Martin Lake 3	595	75	595														
Monticello 3	800	100	800														
Utah Power & Light																	
Hunter 1	360	90	360														
Hunter 2	360	90	360														
Huntington 1	366	85	366														
Naughton 3	330	100	330														
Total	31,738		25,228	4,614	1,896												

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[a] Equivalent scrubbed capacity.

[b] This category includes the flue gas capacity being handled by the FGD system at least part of the time during the report period.

[c] The percent figures listed are average values for all system scrubbing trains during the period.

[d] Flue gas % scrubbed for prototype and demonstration units is not applicable unless the system is designed to bring a unit into compliance with SO_2 emission standard.

[e] Availability, operability, reliability, and utilization as defined in Appendix C of this report.

Table 4. Summary of Operational and Planned Domestic FGD Systems

Company name/ unit name	Capacity MW (gross)	Fuel % sulfur	FGD process	FGD status	System supplier
Alabama Electric					
Tombigbee 2	*255*	*1.15*	*Limestone*	*1*	*Peabody Process Systems*
Tombigbee 3	*255*	*1.15*	*Limestone*	*1*	*Peabody Process Systems*
Arizona Electric Power					
Apache 2	*195*	*0.50*	*Limestone*	*1*	*Research-Cottrell*
Apache 3	*195*	*0.50*	*Limestone*	*1*	*Research-Cottrell*
Arizona Public Service					
Cholla 1	*119*	*0.50*	*Limestone*	*1*	*Research-Cottrell*
Cholla 2	*264*	*0.50*	*Limestone*	*1*	*Research-Cottrell*
Cholla 4	*375*	*0.50*	*Limestone*	*1*	*Research-Cottrell*
Four Corners 1	*175*	*0.75*	*Lime/alkaline flyash*	*1*	*GE Environmental Services*
Four Corners 2	*175*	*0.75*	*Lime/alkaline flyash*	*1*	*GE Environmental Services*
Four Corners 3	*229*	*0.75*	*Lime/alkaline flyash*	*1*	*GE Environmental Services*
Four Corners 4	*755*	*0.75*	*Lime*	*2*	*United Engineers*
Four Corners 5	*755*	*0.75*	*Lime*	*2*	*United Engineers*
Associated Electric					
Thomas Hill 3	*730*	*4.80*	*Limestone*	*2*	*Pullman Kellogg*
Atlantic City Electric					
Cumberland 1	*330*	*3.25*	*Process not selected*	*6*	*Vendor not selected*
Basin Electric Power					
Antelope Valley 1	*440*	*0.68*	*Lime/spray drying*	*2*	*Joy Mfg/Niro Atomizer*
Antelope Valley 2	*440*	*0.68*	*Lime/spray drying*	*2*	*Joy Mfg/Niro Atomizer*
Laramie River 1	*570*	*0.81*	*Limestone*	*1*	*Research-Cottrell*
Laramie River 2	*570*	*0.81*	*Limestone*	*1*	*Research-Cottrell*
Laramie River 3	*570*	*0.54*	*Lime/spray drying*	*2*	*Babcock & Wilcox*
Big Rivers Electric					
D.B. Wilson 1	*440*		*Limestone*	*2*	*Pullman Kellogg*
D.B. Wilson 2	*440*		*Limestone*	*3*	*Pullman Kellogg*
Green 1	*242*	*3.75*	*Lime*	*1*	*American Air Filter*
Green 2	*242*	*3.75*	*Lime*	*1*	*American Air Filter*
Cajun Electric Power					
Chicot 1	*562*	*1.70*	*Process not selected*	*6*	*Vendor not selected*
Chicot 2	*562*	*1.70*	*Process not selected*	*6*	*Vendor not selected*
Chicot 3	*562*	*1.70*	*Process not selected*	*6*	*Vendor not selected*
Chicot 4	*562*	*1.70*	*Process not selected*	*6*	*Vendor not selected*
Oxbow 1	*540*	*0.60*	*Process not selected*	*6*	*Vendor not selected*
Oxbow 2	*540*	*0.60*	*Process not selected*	*6*	*Vendor not selected*
Central Illinois Light					
Duck Creek 1	*416*	*3.66*	*Limestone*	*1*	*Environeering, Riley Stoker*
Duck Creek 2	*450*	*3.30*	*Limestone*	*6*	*Vendor not selected*
Central Illinois Public Service					
Newton 1	*617*	*2.25*	*Dual alkali*	*1*	*GE Environmental Services*
Central Maine Power					
Sears Island 1	*600*	*2.23*	*Process not selected*	*6*	*Vendor not selected*
Central Power & Light					
Coleto Creek 2	*720*	*0.39*	*Process not selected*	*6*	*Vendor not selected*

(continued)

Table 4. (continued)

Company name/ unit name	Capacity MW (gross)	Fuel % sulfur	FGD process	FGD status	System supplier
Cincinnati Gas & Electric					
East Bend 1	650	4.00	Process not selected	6	Vendor not selected
East Bend 2	650	3.00	Lime	1	Babcock & Wilcox
Colorado Ute Electric					
Craig 1	455	0.45	Limestone	1	Peabody Process Systems
Craig 2	455	0.45	Limestone	1	Peabody Process Systems
Craig 3	447	0.45	Lime/spray drying	2	Babcock & Wilcox
Columbus & Southern Ohio Electric					
Conesville 5	411	4.67	Lime	1	Air Correction Division, UOP
Conesville 6	411	4.67	Lime	1	Air Correction Division, UOP
Poston 5	425		Process not selected	6	Vendor not selected
Poston 6	425	2.50	Process not selected	6	Vendor not selected
Commonwealth Edison					
Powerton 51	450	3.53	Limestone	1	Air Correction Division, UOP
Cooperative Power Association					
Coal Creek 1	545	0.63	Lime/alkaline flyash	1	Combustion Engineering
Coal Creek 2	545	0.63	Lime/alkaline flyash	1	Combustion Engineering
Delmarva Power & Light					
Delaware City 1	60	7.00	Wellman Lord	1	Davy McKee
Delaware City 2	60	7.00	Wellman Lord	1	Davy McKee
Delaware City 3	60	7.00	Wellman Lord	1	Davy McKee
Vienna 9	550	2.50	Process not selected	6	Vendor not selected
Deseret Generation & Transmission					
Bonanza 1	410	0.50	Limestone	2	Combustion Engineering
Bonanza 2	410	0.50	Limestone	5	Vendor not selected
Duquesne Light					
Elrama 1-4	510	2.20	Lime	1	GE Environmental Services
Phillips 1-6	408	1.92	Lime	1	GE Environmental Services
East Kentucky Power					
J.K. Smith 1	650	1.50	Lime	4	Babcock & Wilcox
Spurlock 2	500	3.50	Lime	2	Thyssen/CEA
Florida Power & Light					
Martin 3	800		Process not selected	6	Vendor not selected
Martin 4	800		Process not selected	6	Vendor not selected
General Public Utilities					
Coal 1	625	3.50	Process not selected	6	Vendor not selected
Coal 2	625	3.50	Process not selected	6	Vendor not selected
Coal 3	625	3.50	Process not selected	6	Vendor not selected
Coal 4	625	3.50	Process not selected	6	Vendor not selected
Seward 7	690		Process not selected	5	Vendor not selected
Grand Haven Board of Light & Power					
J.B. Sims 3	65	2.75	Lime	2	Babcock & Wilcox

(continued)

Table 4. (continued)

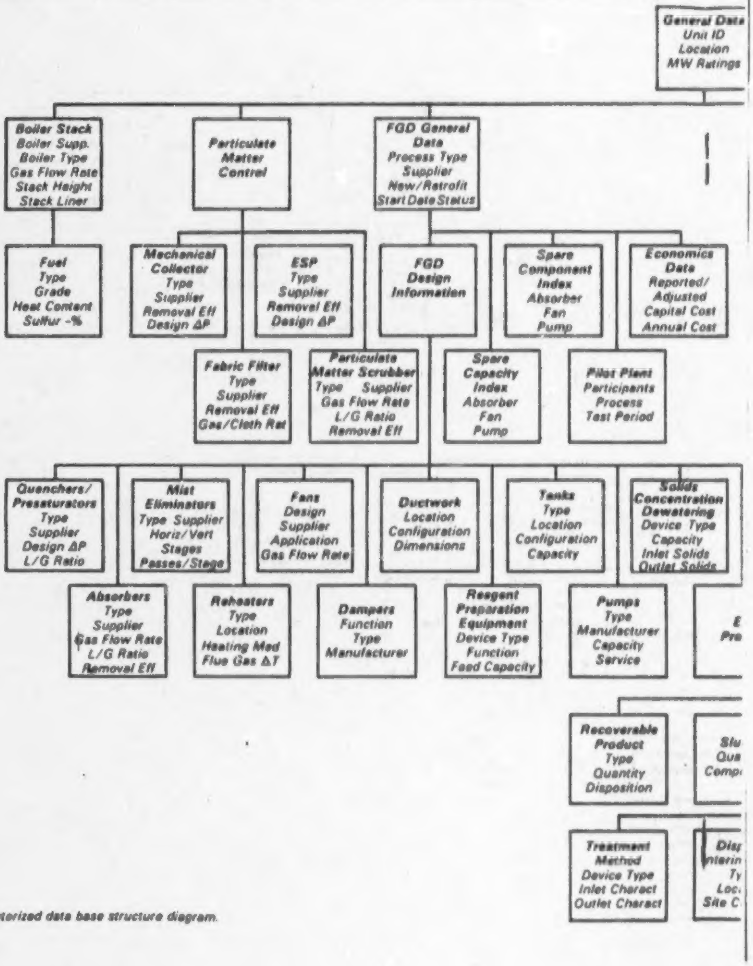
Company name/ unit name	Capacity MW (gross)	Fuel % sulfur	FGD process	FGD status	System supplier
Hoosier Energy					
Merom 1	490	3.50	Limestone	2	Mitsubishi Heavy Industries
Merom 2	490	3.50	Limestone	2	Mitsubishi Heavy Industries
Houston Lighting & Power					
Limestone 1	750	1.08	Limestone	3	Combustion Engineering
Limestone 2	750	1.08	Limestone	3	Combustion Engineering
W.A. Parish 8	600	0.6	Limestone	2	GE Environmental Services
Indianapolis Power & Light					
Patriot 1	650	3.50	Limestone	6	Vendor not selected
Patriot 2	650	3.50	Limestone	6	Vendor not selected
Patriot 3	650	3.50	Limestone	6	Vendor not selected
Petersburg 3	532	3.25	Limestone	1	Air Correction Division, UOP
Petersburg 4	530	3.50	Limestone	2	Research Cottrell
Iowa Electric Light & Power					
Guthrie Co. 1	720	0.40	Limestone	4	Combustion Engineering
Jacksonville Electric Authority					
St. Johns River Power 1	600	2.50	Limestone	5	Vendor not selected
St. Johns River Power 2	600	2.50	Limestone	5	Vendor not selected
Kansas City Power & Light					
Hawthorn 3	90	0.60	Lime	1	Combustion Engineering
Hawthorn 4	90	0.60	Lime	1	Combustion Engineering
La Cygne 1	820	5.39	Limestone	1	Babcock & Wilcox
Kansas Power & Light					
Jeffrey 1	720	0.32	Limestone	1	Combustion Engineering
Jeffrey 2	700	0.30	Limestone	1	Combustion Engineering
Lawrence 4	125	0.55	Limestone	1	Combustion Engineering
Lawrence 5	420	0.55	Limestone	1	Combustion Engineering
Kentucky Utilities					
Green River 1-3	64	4.00	Lime	1	American Air Filter
Hancock 1	708	3.50	Limestone	4	Babcock & Wilcox
Hancock 2	708	3.50	Limestone	6	Vendpr not selected
Lakeland Utilities					
McIntosh 3	364	2.56	Limestone	2	Babcock & Wilcox
Lansing Board of Water and Light					
Erickson 2	160		Process not selected	6	Vendor not selected
Los Angeles Department of Water & Power					
Intermountain 1	820	0.79	Lime	6	Vendor not selected
Intermountain 2	820	0.79	Lime	6	Vendor not selected
Intermountain 3	820	0.79	Lime	6	Vendor not selected
Intermountain 4	820	0.79	Lime	6	Vendor not selected

(continued)

Table 4. (continued)

Company name/ unit name	Capacity MW (gross)	Fuel % sulfur	FGD process	FGD status	System supplier
Louisville Gas & Electric					
Cane Run 4	188	3.75	Lime	1	American Air Filter
Cane Run 5	200	3.75	Lime	1	Combustion Engineering
Cane Run 6	299	4.80	Dual alkali	1	Thyssen/CEA
Mill Creek 1	358	3.75	Limestone	1	Combustion Engineering
Mill Creek 2	350	3.75	Lime	2	Combustion Engineering
Mill Creek 3	427	3.75	Lime	1	American Air Filter
Mill Creek 4	495	3.75	Lime	2	American Air Filter
Paddy's Run 6	72	2.50	Lime	1	Combustion Engineering
Trimble County 1	575	4.00	Process not selected	5	Vendor not selected
Trimble County 2	575	4.00	Process not selected	5	Vendor not selected
Lower Colorado River Authority					
Fayette Power Project 3	435	1.70	Limestone	6	Vendor not selected
Marquette Board of Light and Power					
Shiras 3	44		Lime/spray drying	2	GE Environmental Services
Michigan South Central Power Agency					
Project 1	55	2.25	Limestone	2	Babcock & Wilcox
Middle South Utilities					
Arkansas Lignite 5	890	0.50	Limestone	4	Combustion Engineering
Arkansas Lignite 6	890	0.50	Limestone	4	Combustion Engineering
Unassigned 1	890	0.50	Limestone	4	Combustion Engineering
Unassigned 2	890	0.50	Limestone	4	Combustion Engineering
Wilton 1	890	0.50	Limestone	4	Combustion Engineering
Wilton 2	890	0.50	Limestone	4	Combustion Engineering
Minnesota Power & Light					
Clay Boswell 4	554	0.94	Lime/alkaline flyash	1	Peabody Process Systems
Minnkota Power					
Milton R. Young 2	440	0.70	Lime/alkaline flyash	1	Thyssen/CEA
Monongahela Power					
Pleasants 1	618	3.00	Lime	1	Babcock & Wilcox
Pleasants 2	618	3.00	Lime	1	Babcock & Wilcox
Montana Power					
Colstrip 1	360	0.77	Lime/alkaline flyash	1	Thyssen/CEA
Colstrip 2	360	0.77	Lime/alkaline flyash	1	Thyssen/CEA
Colstrip 3	700	0.77	Lime/alkaline flyash	2	Thyssen/CEA
Colstrip 4	700	0.77	Lime/alkaline flyash	2	Thyssen/CEA
Montana-Dakota Utilities					
Coyote 1	440	0.87	Sodium carbonate/spray drying	1	Wheelabrator-Fry/R.I.
Muscatine Power & Water					
Muscatine 9	166	3.21	Limestone	2	Research-Cottrell

(continued)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Unit ID
Location
MW Ratings
Boiler Stack
Boiler Supp.
Boiler Type
Gas Flow Rate
Stack Height
Stack Liner
Particulate Matter Control
FGD General Data
Process Type
Supplier
New/Retrofit
Start Date Status
Fuel
Type
Grade
Heat Content
Sulfur -%
Mechanical Collector
Type
Supplier
Removal Eff
Design ΔP
ESP
Type
Supplier
Removal Eff
Design ΔP
FGD Design Information
Spare Component Index
Absorber
Fan
Pump
Economics Data
Reported/Adjusted
Capital Cost
Annual Cost
Fabric Filter
Type
Supplier
Removal Eff
Gas/Cloth Rat
Particulate Matter Scrubber
Type Supplier
Gas Flow Rate
L/G Ratio
Removal Eff
Spare Capacity Index
Absorber
Fan
Pump
Pilot Plant
Participants
Process
Test Period
Quenchers/Presaturators
Type
Supplier
Design ΔP
L/G Ratio
Mist Eliminators
Type Supplier
Horiz/Vert
Stages
Passes/Stage
Fans
Design
Supplier
Application
Gas Flow Rate
Ductwork
Location
Configuration
Dimensions
Tanks
Type
Location
Configuration
Capacity
Solids Concentration Dewatering
Device Type
Capacity
Inlet Solids
Outlet Solids
Absorbers
Type
Supplier
Gas Flow Rate
L/G Ratio
Removal Eff
Reheaters
Type
Location
Heating Med
Flue Gas ΔT
Dampers
Function
Type
Manufacturer
Reagent Preparation Equipment
Device Type
Function
Feed Capacity
Pumps
Type
Manufacturer
Capacity
Service
E
Pr
Recoverable Product
Type
Quantity
Disposition
Slu
Qua
Comp
Treatment Method
Device Type
Inlet Charact
Outlet Charact
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nterin
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Figure 1. Computerized data base structure diagram.

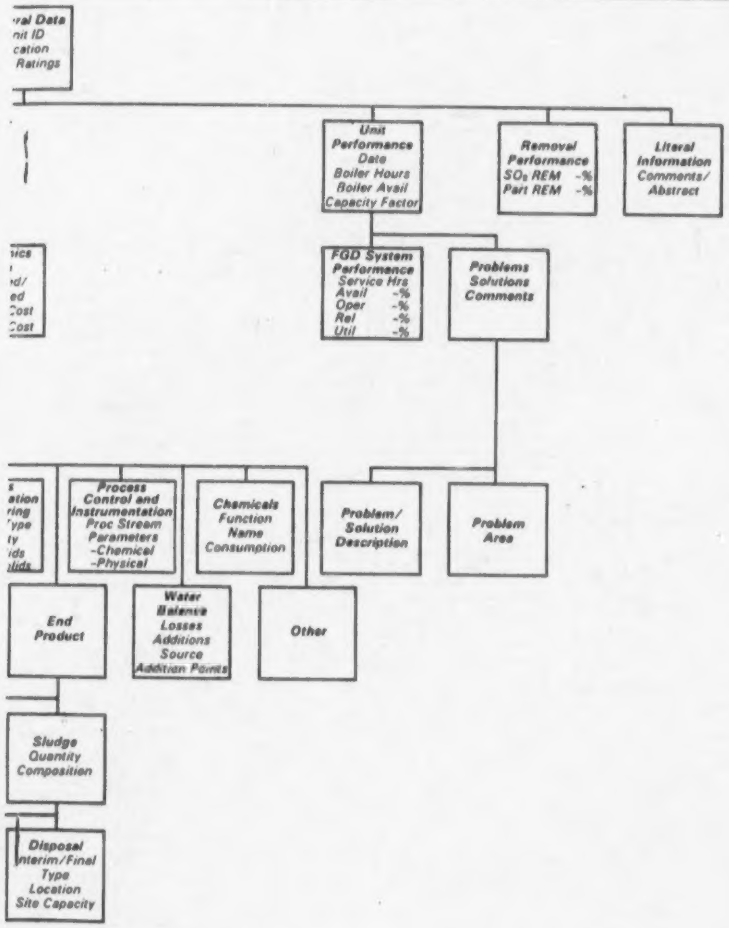
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Table 4. *(continued)*

Company name/ unit name	Capacity MW (gross)	Fuel % sulfur	FGD process	FGD status	System supplier
Nebraska Public Power District					
Fossil III 1	650	0.36	Process not selected	6	Vendor not selected
Nevada Power					
Harry Allen 1	500		Process not selected	6	Vendor not selected
Harry Allen 2	500		Process not selected	6	Vendor not selected
Harry Allen 3	500		Process not selected	6	Vendor not selected
Harry Allen 4	500		Process not selected	6	Vendor not selected
Reid Gardner 1	125	0.50	Sodium carbonate	1	Thyssen/CEA
Reid Gardner 2	125	0.50	Sodium carbonate	1	Thyssen/CEA
Reid Gardner 3	125	0.50	Sodium carbonate	1	Thyssen/CEA
Reid Gardner 4	250	0.75	Sodium carbonate	4	Thyssen/CEA
New York State Electric & Gas					
Somerset 1	625	2.20	Limestone	4	Peabody Process Systems
Niagara Mohawk Power					
Charles R. Huntley 66	100	1.80	Aqueous carbonate/spray drying	2	Rockwell International
Northern Indiana Public Service					
Dean H. Mitchell 11	116	3.50	Wellman Lord	1	Davy McKee
Schahfer 17	421	3.20	Dual alkali	2	FMC
Schahfer 18	421	3.20	Dual alkali	3	FMC
Northern States Power					
Metro Coal 1	200	1.00	Lime	6	Vendor not selected
Riverside 6-7	110	1.20	Lime/spray drying	1	Joy Mfg/Niro Atomizer
Sherburne 1	740	0.80	Limestone/alkaline flyash	1	Combustion Engineering
Sherburne 2	740	0.80	Limestone/alkaline flyash	1	Combustion Engineering
Sherburne 3	860	1.00	Lime	6	Vendor not selected
Pacific Gas & Electric					
Montezuma 1	800	0.80	Limestone	6	Vendor not selected
Montezuma 2	800	0.80	Limestone	6	Vendor not selected
Pacific Power & Light					
Jim Bridger 1	550	0.56	Sodium carbonate	6	Vendor not selected
Jim Bridger 2	550	0.56	Sodium carbonate	6	Vendor not selected
Jim Bridger 2A	550	0.56	Lime/sodium carbonate	2	Fläkt
Jim Bridger 3	550	0.56	Sodium carbonate	6	Vendor not selected
Jim Bridger 4	550	0.56	Sodium carbonate	1	Air Correction Division, UOP
Pennsylvania Power					
Bruce Mansfield 1	917	3.00	Lime	1	GE Environmental Services
Bruce Mansfield 2	917	3.00	Lime	1	GE Environmental Services
Bruce Mansfield 3	917	3.00	Lime	1	Pullman Kellogg
Philadelphia Electric					
Cromby 1	150	3.00	Magnesium oxide	2	United Engineers
Eddystone 1	240	2.60	Magnesium oxide	2	United Engineers
Eddystone 2	334	2.50	Magnesium oxide	2	United Engineers
Plains Electric G&T					
Plains Escalante 1	233	0.80	Limestone	2	Combustion Engineering
Platte River Power Authority					
Rawhide 1	279	0.25	Lime/spray drying	3	Joy Mfg/Niro Atomizer

(continued)

Table 4. (continued)

Company name/ unit name	Capacity MW (gross)	Fuel % sulfur	FGD process	FGD status	System supplier
Public Service Indiana					
Gibson 5	650	3.30	Limestone	2	Pullman Kellogg
Public Service of New Mexico					
San Juan 1	361	0.80	Wellman Lord	1	Davy McKee
San Juan 2	350	0.80	Wellman Lord	1	Davy McKee
San Juan 3	534	0.80	Wellman Lord	1	Davy McKee
San Juan 4	534	0.80	Wellman Lord	2	Davy McKee
Power Authority of State of New York					
Fossil	700	3.00	Process not selected	6	Vendor not selected
Salt River Project					
Coronado 1	350	0.50	Limestone	1	Pullman Kellogg
Coronado 2	350	0.50	Limestone	1	Pullman Kellogg
Coronado 3	400	0.60	Limestone	6	Vendor not selected
San Miguel Electric					
San Miguel 1	400	1.70	Limestone	1	Babcock & Wilcox
Seminole Electric					
Seminole 1	620	2.75	Limestone	2	Peabody Process Systems
Seminole 2	620	2.75	Limestone	3	Peabody Process Systems
Sikeston Board of Municipal Utilities					
Sikeston 1	235	2.80	Limestone	1	Babcock & Wilcox
South Carolina Public Service					
Cross 1	500	1.80	Limestone	3	Peabody Process Systems
Cross 2	500	1.80	Limestone	2	Peabody Process Systems
Winyah 2	280	1.10	Limestone	1	Babcock & Wilcox
Winyah 3	280	1.10	Limestone	1	Babcock & Wilcox
Winyah 4	280	1.70	Limestone	1	American Air Filter
South Mississippi Electric Power					
R.D. Morrow, Sr. 1	200	1.30	Limestone	1	Environeering, Riley Stoker
R.D. Morrow, Sr. 2	200	1.30	Limestone	1	Environeering, Riley Stoker
Southern Illinois Power					
Marion 4	173	3.75	Limestone	1	Babcock & Wilcox
Southern Indiana Gas & Electric					
A.B. Brown 1	265	3.35	Dual alkali	1	FMC
A.B. Brown 2	265	3.35	Process not selected	6	Vendor not selected
Southwestern Electric Power					
Dolet Hills 1	720	0.70	Limestone	3	Air Correction Division, UOP
Dolet Hills 2	720	0.70	Limestone	6	Vendor not selected
Henry W. Pirkey 1	720	0.80	Limestone	3	Air Correction Division, UOP
Soyland Power					
Soyland 1	600	3.00	Process not selected	6	Vendor not selected
Springfield City Utilities					
Southwest 1	194	3.50	Limestone	1	Air Correction Division, UOP

(continued)

Table 4. (continued)

Company name/ unit name	Capacity MW (gross)	Fuel % sulfur	FGD process	FGD status	System supplier
Springfield Water, Light & Power					
Dallman 3	205	3.30	Limestone	1	Research-Cottrell
St. Joe Zinc					
G.F. Weston 1	60	2.00	Citrate	1	Morrison & Knudsen/U.S.B.M.
Sunflower Electric					
Holcomb 1	347	0.47	Lime/spray drying	2	Joy Mfg/Niro Atomizer
Tampa Electric					
Big Bend 4	475	2.35	Lime/limestone	3	Research-Cottrell
Tennessee Valley Authority					
Paradise 1	704	4.20	Limestone	2	GE Environmental Services
Paradise 2	704	4.20	Limestone	2	GE Environmental Services
Shawnee 10A	10	2.90	Lime/limestone	1	Air Correction Division, UOP
Shawnee 10b	10	2.90	Lime/limestone	1	GE Environmental Services
Widows Creek 7	575	3.70	Limestone	1	Combustion Engineering
Widows Creek 8	550	3.70	Limestone	1	Tennessee Valley Authority
Texas Municipal Power Agency					
Gibbons Creek 1	443	1.06	Limestone	2	Combustion Engineering
Texas Power & Light					
Sandow 4	545	1.60	Limestone	1	Combustion Engineering
Twin Oaks 1	750	0.70	Limestone	3	GE Environmental Services
Twin Oaks 2	750	0.70	Limestone	3	GE Environmental Services
Texas Utilities					
Forst Grove 1	750	0.80	Process not selected	5	Vendor not selected
Martin Lake 1	793	0.90	Limestone	1	Research-Cottrell
Martin Lake 2	793	0.90	Limestone	1	Research-Cottrell
Martin Lake 3	793	0.90	Limestone	1	Research-Cottrell
Martin Lake 4	793	0.90	Limestone	3	Research-Cottrell
Mill Creek 1	750		Process not selected	6	Vendor not selected
Mill Creek 2	750		Process not selected	6	Vendor not selected
Monticello 3	800	1.50	Limestone	1	GE Environmental Services
Tucson Electric Power					
Springerville 1	370	0.61	Lime/spray drying	3	Joy Mfg/Niro Atomizer
Springerville 2	370	0.61	Lime/spray drying	3	Joy Mfg/Niro Atomizer
United Power Association					
Stanton 1A	60	0.77	Lime/spray drying	2	Research-Cottrell
Utah Power & Light					
Hunter 1	400	0.55	Lime	1	GE Environmental Services
Hunter 2	400	0.55	Lime	1	GE Environmental Services
Hunter 3	400	0.55	Limestone	2	GE Environmental Services
Hunter 4	400	0.55	Limestone	2	GE Environmental Services
Huntington 1	430	0.55	Lime	1	GE Environmental Services
Naughton 3	330	0.55	Sodium carbonate	1	Air Correction Division, UOP
Washington Water Power					
Creston Coal 1	570		Limestone	6	Vendor not selected
Creston Coal 2	570		Limestone	6	Vendor not selected
Creston Coal 3	570		Limestone	6	Vendor not selected
Creston Coal 4	570		Limestone	6	Vendor not selected

(continued)

Table 4. (continued)

Company name/ unit name	Capacity MW (gross)	Fuel % sulfur	FGD process	FGD status	System supplier
West Penn Power					
Mitchell 33	300	2.80	Lime	2	GE Environmental Services
West Texas Utilities					
Oklaunion 1	720	0.34	Process not selected	3	GE Environmental Services
Oklaunion 2	720	0.34	Process not selected	6	Vendor not selected

1 Operational units.
2 Units under construction.
3 Planned - contract awarded.
4 Planned - letter of intent signed.
5 Planned - requesting/evaluating bids.
6 Planned - considering only FGD systems.

Table 5. Categorical Results of the Reported and Adjusted Capital and Annual Costs for Operational FGD Systems

	Reported						Adjusted					
	Capital			Annual			Capital			Annual		
	Range, $/kW	Average, $/kW	s	Range, mills/kWh	Average, mills/kWh	s	Range, $/kW	Average, $/kW	s	Range, mills/kWh	Average, mills/kWh	s
All	23.7-243.0	79.1	45.4	0.05-13.02	2.64	3.14	35.1-258.9	113.5	54.6	1.69-18.67	7.27	3.82
New	23.7-243.0	77.5	47.2	0.05-5.52	1.84	1.84	35.1-242.1	103.8	43.3	1.69-12.83	6.31	2.68
Retrofit	29.4-157.4	83.2	41.4	0.46-13.02	4.54	4.64	57.5-258.9	132.3	69.4	4.23-18.67	9.12	4.99
Saleable	132.8-185.0	153.0	23.8	13.02-13.02	13.02	0.0	233.6-258-9	249.1	13.6	15.23-18.67	16.44	1.94
Throwaway	23.7-243.0	73.7	41.8	0.05-11.32	2.35	2.65	35.1-242.1	104.2	42.5	1.69-16.27	6.64	3.03
Alkaline flyash/lime	100.3-101.4	101.0	0.6	0.53-2.97	2.16	1.41	133.8-142.9	136.8	5.2	5.75-7.62	6.99	1.08
Alkaline flyash/ limestone	49.3-49.3	49.3	0.0	0.75-0.75	0.75	0.0	94.4-94.4	94.4	0.0	4.99-4.99	4.99	0.0
Dual alkali	47.2-174.8	97.8	67.7	1.30-1.30	1.30	0.0	80.6-242.1	134.6	93.1	4.59-12.83	7.79	4.41
Lime	29.4-243.0	80.1	50.6	0.92-11.32	3.55	3.02	57.5-192.7	105.1	39.1	3.70-16.27	7.79	3.63
Limestone	23.7-168.0	65.0	34.9	0.05-7.76	1.78	2.54	35.1-148.7	93.4	42.5	1.69-10.44	5.61	3.65
Sodium carbonate	42.9-100.8	69.2	30.7	0.23-0.46	0.38	0.13	79.9-138.5	101.7	28.0	5.29-6.78	5.88	0.73
Wellman Lord	132.8-185.0	153.1	23.8	13.02-13.02	13.02	0.0	233.6-258.9	249.1	13.6	15.23-18.67	16.44	1.94

Current projections indicate that the total power generating capacity of the U.S. electric utility industry will be approximately 831 GW by the end of 1999.[1] (This value reflects the annual loss resulting from the retirement of older units, which is considered to be 0.4% of the average generating capacity at the end of each year.[2]) Approximately 373 GW or 45% of the 1999 total will come from coal-fired units. The distribution of power generation sources, both present (December 1980) and future (December 1999) is shown in Table 6.[1]

In light of the revised New Source Performance Standards, actual FGD control is expected to be greater than that reflected by the figures above. For example, about 50 to 60 systems representing approximately 29,000 to 31,000 MW of generating capacity presently fall into the uncommitted category. These are systems that cannot be included in the committed group at this time because information regarding their status is not ready for public release.

In an effort to show general FGD usage trends, Table 8 gives current (September 1981) and projected (December 1999) breakdowns of throwaway product systems versus saleable product systems as a percent of the total known commitments to FGD as of the end of the third quarter 1981.

Highlights: July-September 1981

The following paragraphs highlight FGD system developments during the third quarter 1981.

Arizona Electric Powdr reported that the Apache 2 FGD system achieved 100% availability for July, August, and September. The Apache 3 FGD system achieved 100% availability during July and August but dropped to 62% availability during September as a result of damper problems.

Arizona Public Service announced that construction began during the period on the lime FGD systems that are being retrofitted to control SO_2 emissions from Four Corners 4 and 5. Both units are rated at 755 MW (gross) and burn coal with an average sulfur content of 0.75%. The FGD systems are being supplied by United engineers and are slated for initial start-up in December 1984.

Basin Electric Power announced that construction of the lime/spray drying FGD system being installed on Antelope Valley 2 began during the third quarter.

Table 6. Power Generation Sources: Present and Future

	Coal	Nuclear	Oil	Hydro	Gas	Other	GW (total)
December 1980	41%	10%	24%	12%	12%	1%	616
December 1999	45%	15%	19%	11%	9%	1%	831

Based on the known commitments to FGD by utilities as presented in Table 1, the percentage of electrical generating capacity controlled by FGD for both the present (September 1981) and the future (December 1999) is shown in Table 7.

Table 7. FGD Controlled Generating Capacity: Present and Future

	Coal-fired generating capacity controlled by FGD, %	Total generating capacity controlled by FGD, %
September 1981*	13.8	8.7
December 1999	28.7	12.9

** The number of committed FGD systems is as of September 1981; however, the figure used for the total generating capacity and coal-fired generating capacity is based on the available December 1980 figures.*

Table 8. Summary of FGD Systems by Process

		Percent of total MW	
		September 1981	December 1999
Throwaway product process			
•Wet systems			
Lime		36.8	20.6
Limestone		48.7	36.9
Dual alkali		3.7	2.0
Sodium carbonate		4.0	3.1
NA[a]		–	5.2
•Dry systems			
Lime		0.3	3.4
Lime/sodium carbonate		–	0.1
Sodium carbonate		1.4	0.4
Saleable product process			
•Process	•Byproduct		
Aqueous carbonate/ spray drying	Elemental sulfur	–	0.1
Citrate	Elemental sulfur	0.2	0.1
Lime	Gypsum	–	0.1
Limestone	Gypsum	–	0.2
Lime/limestone	Gypsum	–	0.5
Magnesium oxide	Sulfuric acid	–	0.7
Wellman Lord	Sulfuric acid	2.3	1.2
Wellman Lord	Elemental sulfur	2.6	0.8
Process undecided		–	24.6
Total		100.0	100.0

[a] NA - Not available (these systems are committed to a throwaway product process; however, the actual process is unknown at this time).

This new unit will be rated at 440 MW (gross) and will fire coal with an average sulfur content of 0.68%. The FGD system is being supplied by Joy Manufacturing/Niro Atomizer and is scheduled for initial operation in October 1985.

The Laramie River 1 FGD system achieved availabilities of 100%, 97%, and 100% during July, August, and September, respectively. Corrosion encountered in the modular quencher sections necessitated repairs during the period; however, overall system availability remained high as a result of spare module capacity.

Initial operations of the Laramie River 2 FGD system of Basin Electric Power began during July. Laramie River 2 is rated at 570 MW (gross) and fires coal with an average sulfur content of 0.81%. The limestone FGD system, which was supplied by Research-Cottrell, consists of five absorber modules and is preceded by an ESP for primary particulate matter removal. Availability of the FGD system during August and September exceeded 95%.

Big Rivers Electric reported that the Green 1 FGD system achieved availabilities of 98%, 95%, and 85% during July, August, and September, respectively. The availability dropped slightly in September due to the replacement of the chevron mist eliminators in both modules. The Green 2 FGD system achieved availabilities of 96%, 97%, and 100% for the same 3 months with only minor problems being encountered.

Plans for the installation of Unit 2 at the Duck Creek Generating Station have been delayed as a result of current load projections and economic conditions. As a result, the utility has temporarily suspended requesting/evaluating bids for an FGD system.

Columbus and Southern Ohio Electric reported availabilities of 100%, 99%, and 94% for the Conesville 5 FGD system during July, August, and September, respectively. Conesville 6 achieved availabilities of 99%, 99%, and 95% during the same 3 months. Both units were removed from service on September 19 as a result of low power demand.

Deseret Generation and Transmission announced during the third quarter that Moon Lake 1 and 2 have been renamed. The units are now Bonanza 1 and 2.

East Kentucky Power announced that plans for the installation of J.K. Smith 2 have been postponed indefinitely as a result of power demand not meeting utility projections. In addition, initial start-up of

J.K. Smith 1 has been pushed back to August 1987.

A letter of intent was signed during the period by Kentucky Utilities with Babcock and Wilcox for the installation of a limestone FGD system to control SO_2 emissions from the planned Hancock 1. This new unit is to be rated at 708 MW (gross) and will fire coal with an average sulfur content of 3.50%. The FGD system is scheduled to begin initial operations in April 1989.

Plans for the installation of Unit 2 at the Hancock Generating Station have been delayed and Kentucky Utilities has temporarily suspended requested/evaluating bids for the installation of an FGD system for SO_2 control. The unit, which is to be rated at 708 MW (gross), is currently scheduled for operation in 1994.

Nevada Power announced during the period that plans for Warner Valley 1 and 2 have been postponed indefinitely. Due to projected SO_2 emissions and the planned station location with respect to a Class I area in Utah, the utility has been unable to obtain a permit from the EPA. The units were to be rated at 295 MW (gross) and were slated for operation in June 1985 and June 1986, respectively.

Plans for the installation of Wisconsin Coal 1 by Northern States Power have been postponed indefinitely. The unit was to be rated at 670 MW (gross) and was slated for initial operations in 1981.

Pacific Power and Light has temporarily suspended requesting/evaluating bids for the installation of FGD systems at the planned Jim Bridger Units 1, 2, and 3 as a result of obtaining a modification in the compliance deadline for Unit 3. The units are to be rated at 550 MW (gross) each and are scheduled for operation in 1988, 1990, and 1986, respectively.

Construction began during the third quarter 1981 on the limestone FGD system being installed on Plains Escalante 1 of Plains Electric Generation and Transmission. This new unit will be rated at 233 MW (gross) and will fire coal with an average sulfur content of 0.80%. The FGD system, which is being supplied by Combustion Engineering, is slated for initial operation in December 1983.

Initial operations of the limestone FGD system installed on San Miguel 1 of San Miguel Electric began in August 1981 when one of the four absorber modules was placed in service. San Miguel 1 is rated at 400 MW (gross) and fires coal with an average sulfur content of 1.70%. The FGD system, which was supplied by Babcock and Wilcox, is preceded by a cold-side ESP for primary particulate matter control. Initial reports indicate that boiler problems have limited operation of the FGD system.

Construction of the Seminole 1 limestone FGD system of Seminole Electric began during the period. The 620 MW (gross) new unit will fire coal with an average sulfur content of 2.75%. The FGD system is being supplied by Peabody Process Systems and will be preceded by ESPs for primary particulate matter removal. Initial operation is scheduled for March 1983.

Initial operations of the American Air Filter limestone FGD system installed on Winyah 4 began in July when one module was placed in service. Winyah 4 is rated at 280 MW (gross) and fires coal with an average sulfur content of 1.70%. A fire in the second module 2 months prior to projected start-up has kept that module off line. Integrated operation of both modules is expected to begin in January or February 1982.

South Mississippi Electric Power reported availabilities of 100%, 100%, and 99% for the R.D. Morrow 1 FGD system during July, August, and September, respectively. The R.D. Morrow 2 FGD system achieved availabilities of 100%, 100%, and 98% during the same 3 months. No major FGD related problems were encountered during the third quarter.

Southern Illinois Power has announced that plans for the installation of Marion 5 have been postponed indefinitely as a result of current load projections, economic conditions, and environmental concerns. The unit was to have been rated at 300 MW (gross) and was slated for operation in 1988.

The A.B. Brown 1 FGD system of Southern Indiana Gas and Electric achieved availabilities of 91%, 92%, and 98% during July, August, and September, respectively. Some pump liner problems were noted during the 3 months.

Construction of the lime/spray drying FGD system being installed on Holcomb 1 of Sunflower Electric began during the third quarter. Holcomb 1 is to be rated at 347 MW (gross) and will fire coal with an average sulfur content of 0.47%. The FGD system, which is being supplied by Joy Manufacturing/Niro Atomizer, is scheduled to begin initial operations in September 1983.

Initial operations of the sodium carbonate FGD system retrofitted on Naughton 3 of Utah Power and Light began in early September. Naughton 3 is rated at 330 MW (gross) and fires coal with an average sulfur content of 0.55%. The FGD system was supplied by the Air Correction Division, UOP and is preceded by an ESP for primary particulate matter control. The FGD system is currently in the start-up phase of operations and is not required to be in full service until the end of the year.

West Texas Utilities announced that a contract has been awarded to General Electric Environmental Services for the installation of a limestone FGD system to control SO_2 emissions from Oklaunion 1. This new unit will be rated at 720 MW (gross) and will fire coal with an average sulfur content of 0.34%. FGD system start-up is scheduled for December 1986.

References

1. U.S. Department of Energy. Energy Information Administration. Office of Coal and Electric Power Statistics. Electric Power Statistics Division. *Inventory of Power Plants in the United States, 1980 Annual.* Publ. No. DOE/EIA-0095 (80).
2. Berman, Ira M. *New Generating Capacity: When, Where, and by Whom.* Power Engineering 85(4)72. April 1981.

The first commercial dry SO2 scrubber is discussed in this Chemical Engineering Progress article. Successfully operating commercially with high SO2 removal, dry FGD offers advantages over wet FGD. Dry FGD is also of special interest because it can be used on commercial boilers which are generally smaller than utility boilers but contribute significantly to man-made sulfur oxide pollution.

REDUCING PLANT POLLUTION EXPOSURE:

Dry FGD on an Industrial Boiler

The first commercial dry flue gas desulfurization system, installed on an industrial pulverized coal-fed boiler, has been operating for more than 20,000 hours with an availability of more than 97%.

T. V. Reinauer, MikroPul Corp., Summit, N.J. 07901
J. P. Monat, Abcor, Inc., Wilmington, Mass. 01887
M. Mutsakis, Koch Engineering Co., Inc., New York, N.Y. 10017

In September of 1978, the Hammermill Paper Co. of Erie, Pa., awarded the MikroPul Corp. a contract to install a dry flue gas desulfurization (FGD) system. This system was to remove SO_2 and particulates from the flue gas stream of a Riley Stoker Corp.'s pulverized coal-fired boiler at the Strathmore Paper Co. in Woronoco, Mass. The boiler was originally designed to burn coal but was subsequently converted to oil. The boiler was reconverted to coal-firing in 1979.

It was required that the system, Figure 1, comply with the Commonwealth of Massachusetts emission standards for the Pioneer Valley District while firing coal of sulfur contents up to 3.0%. This requirement includes both SO_2 and particulate regulations which specify a maximum of 5.16×10^{-7} kg/kJ (1.2 lb/million Btu) of SO_2 and maximum of 4.30×10^{-8} kg/kJ (0.10 lb/million Btu) of particulates. These levels are equivalent to the federal new source performance standards.

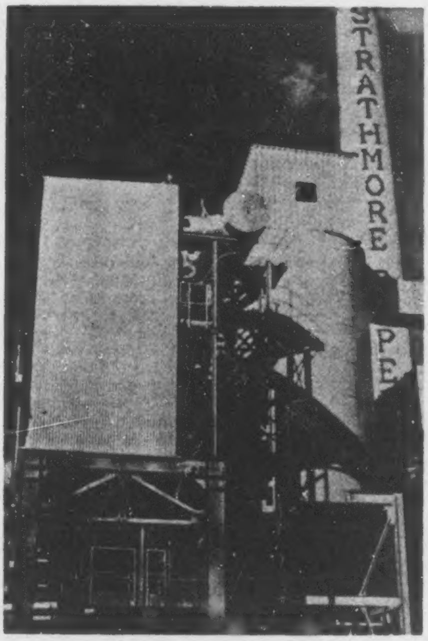


Figure 1. First commercial dry SO_2 scrubbing system.

0360-7275/83/6583-0074 $02.00 © 1983 AIChE

As of the spring of 1982 the system had logged in excess of 20,000 operating hours.

Process

The Strathmore dry FGD system consists of two principal items of hardware: the spray dryer reactor and a fabric filter.

A typical dry FGD scrubbing system is shown in Figure 2. Figure 3 illustrates the process flow. Bulk deliveries of quicklime (CaO) are pneumatically conveyed to a lime storage silo. From here, the lime is fed to a lime slaker where it is hydrated to $Ca(OH)_2$. Residual grit is removed by a vibrating screen as the hydrated lime is transferred to the agitated slurry holding tank. A pump delivers the lime slurry at a controlled rate and concentration to two-fluid nozzle atomizers on top of the spray dryer. In the spray dryer, the atomizers create a spray of fine droplets, which are intimately mixed with the hot incoming flue gas.

The sulfur oxides contained in the flue gas are absorbed by the fine droplets and react with the lime alkali to form insoluble calcium sulfite and sulfate, Table 1. At the same time, the hot flue gas evaporates the water in the droplets to produce an insoluble dry powder. The evaporation of water cools the flue gas to a temperature of 11 to 22 K (20 to 40°F) above the dew point of 322 to 328 K (120 to 130°F). The powder leaving the spray dryer consists of the reaction products, fly ash, and unreacted lime. The scrubbed flue gas and suspended fly ash and powder are separated by the downstream fabric filter.

Reagent Preparation. Soda ash and lime are the only two reagents that have been proven commercially feasible to date. Lime is the reagent chosen for most dry scrubbers because of its availability, low cost, and the stability of its reaction products. Lime preparation is more complicated than soda ash preparation since lime must be slaked (hydrated) from the delivered form of quicklime (CaO) to form a hydrated lime slurry ($Ca(OH)_2$). The technique applied in slaking lime can significantly affect its reactivity. Therefore, the slaking method is one of the most important design considerations.

The water used for slaking must be of reasonably good quality. Ideally, it should be softened to prevent scaling problems and must be low in sulfates and other chemicals that cause a reduction in reactivity.

It has been found that, in certain cases, recycle of some of the collected product from the fabric filter can improve the overall lime utilization. This recycling permits the reuse of

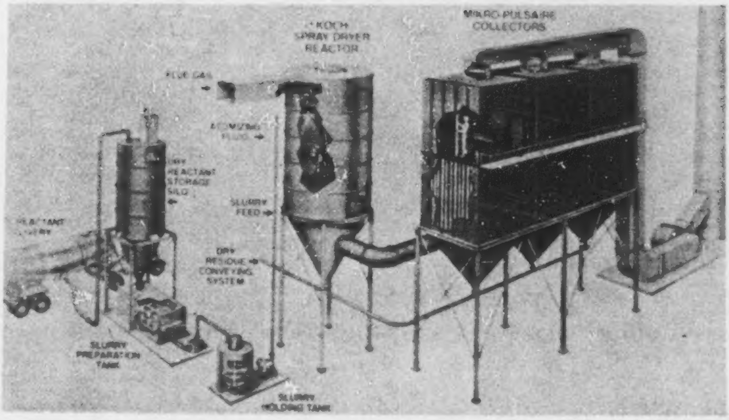
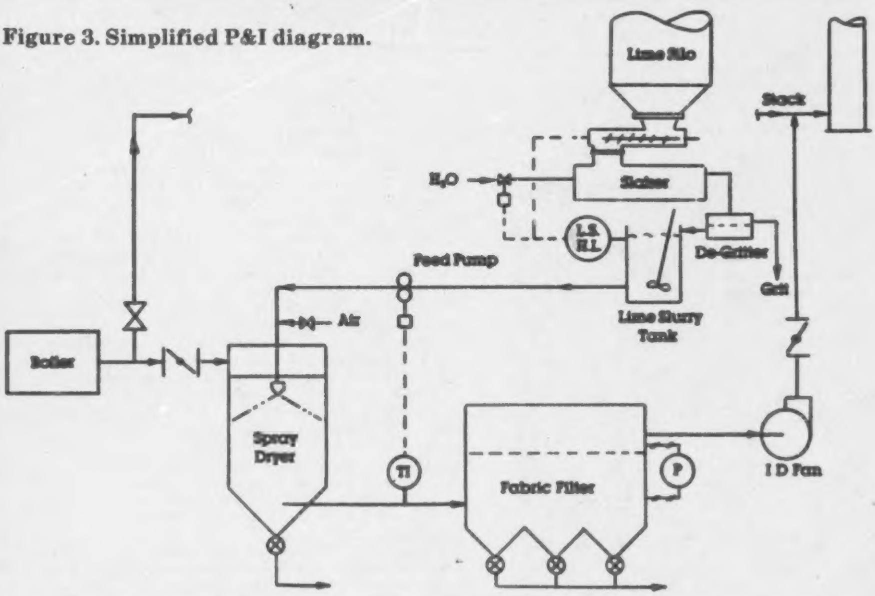

KOCH SPRAY DRYER REACTOR
MIKRO-PULSAIRE COLLECTORS
FLUE GAS
ATOMIZING FLUID
DRY REACTANT STORAGE SILO
SLURRY FEED
REACTANT DELIVERY
DRY RESIDUE CONVEYING SYSTEM
SLURRY PREPARATION TANK
SLURRY HOLDING TANK

Figure 2. Typical dry FGD scrubbing system.

Figure 3. Simplified P&I diagram.


Lime Silo
Stack
H_2O
Slaker
L.S.
H.L.
De-Gritter
Grit
Feed Pump
Lime Slurry Tank
Air
Boiler
Spray Dryer
TI
Fabric Filter
P
I D Fan

unreacted lime in the spent product; if the fly ash is high in alkalinity, it may reduce the required amount of lime reagent. This technique has not been evaluated at Strathmore.

The amount of pumped slurry required falls in the range of 26.7 to 53.5 cm^3/m^3 (0.2 to 0.4 gal/1,000 ft^3) of flue gas. Wet scrubber slurry pumping rates are approximately 200 times this rate.

The *boiler* consists of a pulverized coal and oil-fired unit rated for coal at 8.2 kg/s (65,000 lb/h) of steam. The cogeneration boiler provides heating and process steam to the paper mill and drives a 3.2-MW generator; excess power is sold to the local electric utility.

The *spray dryer* is constructed of stainless steel. It is 4.4 m (14.5 ft) in diameter and has an overall side height of 8.5 m (28 ft). The flue gas enters a plenum at the top of the dryer and is distributed equally to four diffusers into which individual nozzle lance assemblies are inserted. The nozzles are of the two-fluid external mix type. Atomization is accomplished with compressed air.

Table 1. Chemistry of SO_2 absorption.

Lime Slaking
$CaO(S) + H_2O \rightarrow Ca(OH)_2(S) \rightleftharpoons Ca^{++} + 2OH^-$

Reaction with Hydroxide
$SO_2(L) + OH^- \rightleftharpoons HSO_3^-$
$HSO_3 + OH^- \rightleftharpoons SO_3^= + H_2O$

Precipitation of Products
$SO_3^= + Ca^{++} \rightleftharpoons CaSO_3(S)$
$SO_4^= + Ca^{++} \rightleftharpoons CaSO_4(S)$

SO_2 Absorption
$SO_2(G) \rightleftharpoons SO_2(L)$

Partial Oxidation of Sulfite
$SO_3^= + \frac{1}{2}O_2(L) \rightleftharpoons SO_4^=$

Overall Reaction

$$SO_2(G) + Ca(OH)_2(S) \xrightarrow{H_2O} CaSO_3(S) + H_2O$$

$$SO_2(G) + Ca(OH)_2(S) + \tfrac{1}{2}O_2(G) \xrightarrow{H_2O} CaSO_4(S) + H_2O$$

The flue gas exits horizontally from the dryer cone. The exit is designed with the intent to carry over the maximum possible quantity of powder to the fabric filter.

The dryer controls, Figure 4, are basic, but include a number of safety features. In case of failure of lime supply, the system is switched to water (which is also applied during shutdown to purge the system of residual lime). The slurry and water piping is insulated and strip heated to prevent freeze-up during cold weather.

The *fabric filter* system consists of three top bag access pulse-jet collectors in parallel, with manually operated isolation dampers so that any single unit may be taken off line for service. Based on 18.9 m³/s (40,000 ft³/min), the filtration velocity is 20 mm/s (3.9 ft/min) with three units on line and 29 mm/s (5.8 ft/min) with one unit off line. Because of the temperature reduction due to evaporation, the filtration velocity is actually somewhat lower. The hoppers and tubesheets are constructed of stainless steel, while the balance of the unit is constructed of mild steel with a coal tar epoxy coating. The filter medium is homopolymer acrylic felt with a special treatment finish. The bag retainers are constructed of mild steel with high-temperature epoxy coating.

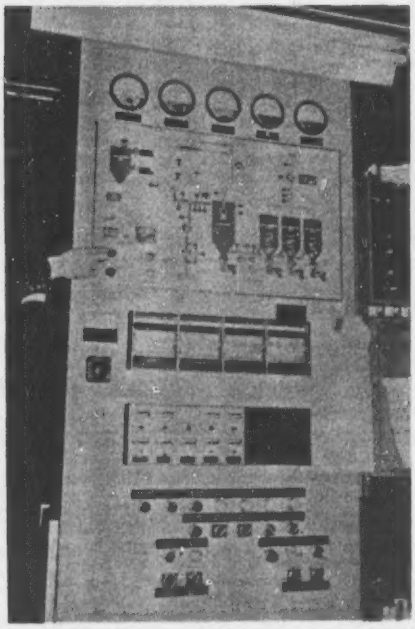


Figure 4. Dry FGD control panel.

The *lime slaker* is a detention-type slaker constructed of mild steel and includes an integral grit remover as well as the necessary controls. It has a capacity for slaking 0.315 kg/s (2,500 lb/h) of quicklime.

A 0.61 m (24 in.) vibrating screen is used to remove residual grit from the slurry delivered from the slaker. The object is to avoid grit of a particle size larger than 40 mesh from passing into the lime system.

Insulation. In consideration of the severity of the winter weather of western Massachusetts, the entire system from the I.D. fan outlet to the stack is insulated with 0.15 m (6 in.) of insulation and aluminum external lagging.

The SO_2 *monitoring instrument* is an ultra-violet (UV) photometric extractive SO_2 analyzer and recorder. Probes are installed at three points: dryer inlet, dryer exit, and fabric filter outlet. A strip chart continuously records the SO_2 ppm at the three points. Although it is necessary to monitor fabric filter exit loadings only, the three-point system provides a means of measuring the efficiency of each component of the system independently; i.e., the SO_2 removal in the spray dryer itself, and the removal in the fabric filter.

The following key items are also recorded: slurry flow rate, dryer inlet temperature, dryer outlet temperature, and Pulsaire Collector temperature. An annubar is located in the fabric filter outlet duct to measure the gas flow through the system. Various system process operating data, Figure 5, are logged on an hourly or daily basis.

System performance

Since startup, the sulfur content of the coal has varied dramatically, from a low of near compliance levels to 5% sulfur. However, the capability of the system to reach compliance levels of SO_2 emissions when handling 5% sulfur coal was also demonstrated. The heating value of the coal generally ranges from 29.1 to 31.5 kJ/g (12,500 to 13,500 Btu/lb). The present coal specifications call for 3% sulfur.

The system normally operates 24 hours a day throughout the year. Although the boiler is usually operated at 32 Mg (70,000 lb) of steam per hour, it is common to see sudden and wide boiler load fluctuations due to changes in process steam demands. Figure 6 shows a three-hour time increment which indicates typical steam demand conditions. Common load demand variations are depicted in Figure 7. These load changes have little effect on the FGD system because of automatic controls.

The fabric filter system has delivered outstanding performance. In spite of numerous upsets in the spray dryer, the bags have never become wet. The differential pressure has always been controllable in the range of 0.5 to 1.0 kPa (2 to 4 in.) at a pulse cleaning cycle of about seven minutes. When one unit is isolated, reducing the cleaning cycle to three minutes maintains the desired differential pressure. There has never been a problem of moisture condensation or corrosion anywhere in the dust collector chamber, even during many days of subzero weather.

The bag set that was in use at Strathmore, as of May 1, 1982, had been in service for more than 30 months while maintaining a particulate discharge of not more than 9.2 mg/m^3 (0.004 gr/ft^3). Strathmore executed a complete bag replacement in June 1982.

During the early stages of Strathmore system operation, it became apparent that the FGD system as initially designed had some serious shortcomings. These included frequent wetting-out of the dryer walls, insufficient SO_2 removal, poor atomization, and nozzle abrasion. Although some of these problems arose because of faulty design, many of them were due to inaccurate specifications. For example, the volumetric air flow through the dryer was 25–35% greater than specified, because of in-leakage downstream of the combustion chamber. This caused a reduced dryer residence time and contributed to wet-out.

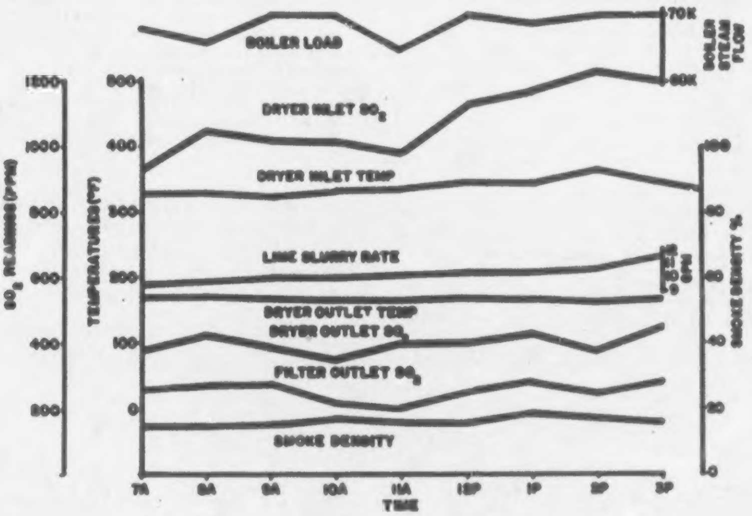
It was quickly noted that the original two-fluid nozzles were inadequate; the atomization was poor and the abrasion rates were extremely high. The nozzles were replaced with technologically improved, two-fluid external mix nozzles. The result was dramatically favorable in terms of proper atomization and the elimination of abrasion. However, wetting-out continued (although less frequently) and there was still a deficiency in SO_2 removal.

To solve those problems, MikroPul consulted with Koch Engineering Co. in late 1979. Koch Engineering recommended effective, low-cost system modifications. These included: changes in the spray dryer aerodynamics to improve flue gas/slurry mixing; changes in the slurry feed and distribution system; and changes in operating procedure. In March of 1980, MikroPul modified the lime system and installed the proprietary Koch spray dryer internals. Startup and operation of the spray reactor were subsequently guided by Koch's high-technology unit, Abcor, Inc.

After the modifications were completed, the system proved capable of meeting performance levels above and beyond those specified in the contract.

The use of four-nozzle atomizers within one dryer was found to be a definite advantage over rotary disc atomizer setups, wherein only one atomizer is used per dryer. The arrangement of applying four separate spray nozzles in the dryer allows for on-line inspection and/or maintenance of nozzles without interruption of the performance of the system.

It takes an operator only a few minutes to remove a nozzle lance assembly and place a spare assembly into operation. When a nozzle is removed, the SO_2 outlet concentration rises slowly, but the automatic dryer controls quickly supply additional slurry to the remaining three nozzles, causing SO_2 removal to stabilize at its previous level. Furthermore,



SI Conversion: °C = (°F − 32)/1.8; L/s = gpm × 0.0631

Figure 5. Daily log entries for emission control.

the two-fluid nozzle system provides a means for unusually high on-line performance. During the 90-day demonstration period, the system was on stream 99% of the scheduled hours of operation.

The *slaker* problems primarily related to the development of operator techniques and training, as well as to equipment adjustment for slaking various types of lime. The various limes tested included high-calcium pebble lime, pulverized lime, granulated lime, and dolomitic lime. Also, some attempts were made to use waste lime products of high alkalinity.

We were hopeful that the dolomitic lime would have performed acceptably, since it is much less expensive and more readily available. Although the laboratory results on dolomitic lime were encouraging, the field results were disappointing when high slurry concentrations were required for high-sulfur coal. High-calcium pebble lime proved to be best for this particular case, demonstrating that care should be applied to the selection and quality control of the lime used for dry scrubbing.

The *waste products* are deposited in portable containers that have been adapted to receive material directly from the dust collectors. The containers are sealed so that powder does not escape during transit to the landfill sites. The rate of buildup in the containers depends on the ash and sulfur content of the coal and the amount of lime sprayed into the dryer. Under peak demand conditions, it is necessary to remove one of the containers every 18 hours.

For the year 1981, 20,523 metric ton (Tm) (22,600 ton) of coal was burned. Therefore, approximately 2,360 Tm (2,600 ton) of fly ash combined with nearly 3,800 Tm (4,200 ton) of additional waste material was generated by the FGD system.

Throughout the period of operations that commenced in August of 1979, the data on SO_2 were taken with the UV photometric analyzer which has performed extremely well. To confirm the SO_2 analyzer's performance, five extensive EPA (Environmental Protection Agency) Method Six tests were conducted at spaced time intervals by Mostardi-Platt to compare results. In all cases, the maximum deviation was 15%. The good agreement between the Mostardi-Platt tests and the SO_2 analyzer corroborates the following test results.

Compliance tests

The system compliance tests were performed by Mostardi-Platt Assoc. on March 25 and 26, 1981. R. Wineberg of the Massachusetts Department of Environmental Quality observed the tests, during which fuel samples were taken for analysis to obtain the data required for emission rate calculations. Sample increments were taken every 15 minutes. The "as received" coal had a sulfur level of 3.79% and a heating value of 30.07 kJ/g (12,905 Btu/lb), while on a dry basis the coal was 3.95% sulfur and 31.35 kJ/g (13,456 Btu/lb). During the tests, 94.24% of the energy input derived from coal and 5.76% from oil; the boiler was operated at a steady load of approximately 8.23 kg/s (65,300 lb/h) of steam.

During this test program, the average particulate emission rate, based on three test runs, was 6.4 ng/J (0.015 lb/10^6 Btu) at an average flue gas volumetric flow rate of 16.2 m^3/s (34,312 actual ft^3/min). This is one-seventh of the allowable limit. In terms of particulate exit loading, the rate was 9.2 mg/m^3 (0.004 gr/ft^3).

The average SO_2 emission rate as determined in the outlet duct was 176 ng/J (0.410 lb/10^6 Btu). All emission rates were determined with F-factors that were calculated from flue gas composition. Flue gas analyses were obtained with an Orsat analyzer. The SO_2 removal efficiency of the system ranged from 90.1 to 96.6% with an average efficiency over six tests of 92.4%. Emission results are:

	As Tested	Allowable Limit
Particulate	6.4 ng/J (0.015 lb/10^6 Btu)	43 (0.10)
SO_2	176 ng/J (0.410 lb/10^6 Btu)	520 (1.2)

Records have been compiled of costs incurred by burning coal (instead of oil) at the Strathmore Paper Co. These include labor, maintenance, power, reactant and disposal. The costs amount to $16.52/metric ton ($15/ton) of coal burned. Over most of this time period, the cost of the rail delivery of coal has been $45.25/metric ton ($42/ton).

The cost of Number 6 compliance fuel oil (1% sulfur) on an equivalent Btu basis was (in March, 1981) approximately $159.69/metric ton ($145/ton). The savings are, therefore, $96.92/metric ton ($88/ton) of coal burned. So, for the typical case of 75 ton/d of coal, the daily net savings are about $6,000, or 61% of the fuel cost; this was made possible by the dry FGD system, Table 2. Operating horsepower of combined coal-fired boiler and dry FGD system is:

Figure 6. Normal steam flow (A), and steam temperature (B), consistent with normal boiler load demand.

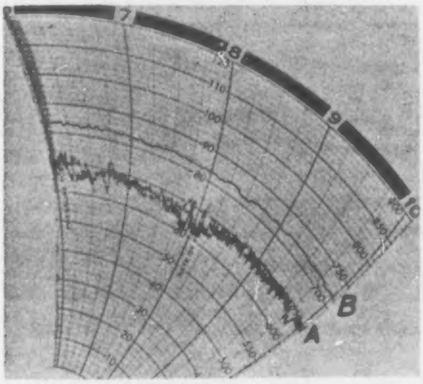


Figure 7. Steam flow (A), air flow (B), and steam temperature (C) illustrate how rapidly changing steam flow demand is readily met over wide load ranges.

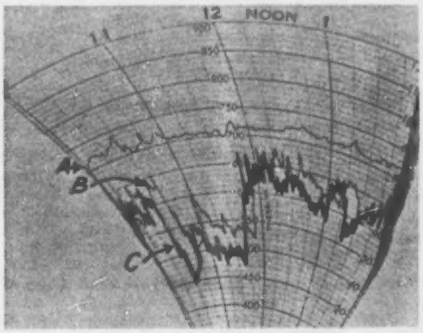


Table 2. Economic justification of dry FGD system.

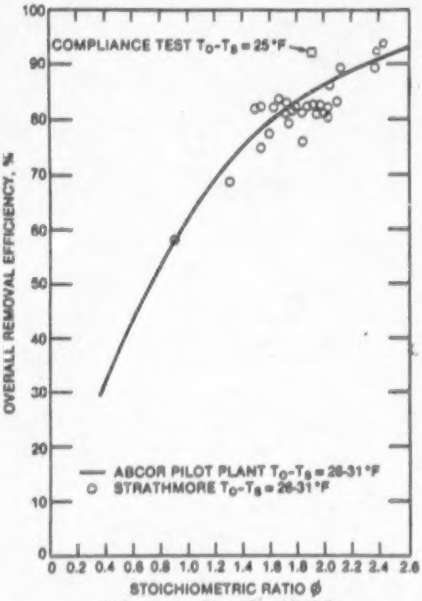
Coal Burned (22,600 ton/yr)		
Cost of Coal at Mine		$35.50/ton
Freight Cost		$22.69
Delivered Cost		$58.19/ton
Cost Associated with Normal Coal Use		
Labor Cost (Receiving, Handling, Storage, etc.)		$ 0.05/ton
Unloading Ash Silo and Trucking to Dump		0.80
Dump Fees for Fly Ash		0.48
Miscellaneous (Coal Testing, Frozen Coal, Clean Yard, etc.)		0.35
	Subtotal:	$ 2.68/ton
Cost Associated with Dry Scrubber System		
Lime Delivered by Truck to Mill		$ 8.89/ton
Baghouse Refuse Handling		2.50
Dumping Fees for Scrubber Waste		0.41
Lime Slurry Additives (Antifloculant)		0.24
Outside Maintenance Services		0.37
Additional Power Plant Operating Crew (3 Men)		2.65
	Subtotal:	$16.06/ton
Total Cost per ton of Coal		$76.93/ton
Oil Equivalent of Coal Burned (92,660 bbl/yr)		
Oil Cost (1981 Average)		$33.36/bbl
Annual Cost of Oil Equivalent of Coal		$3,118,900
Annual Cost for Coal and Dry Scrubber		$1,738,600
Savings per Year		$1,380,300

SI Conversion: L = bbl × 159

Boiler Related: 137 kW (183 hp), includes F.D. fan, I.D. fan, pulverizer, coal feeder

FGD Related: 187 kW (251 hp), includes booster I.D. fan, compressors, airlocks, pumps, agitators, slaker

Total: 324 kW (434 hp)



SI Conversion: °C = (°F − 32)/1.8

Figure 8. Stoichiometric ratio vs. SO_2 removal.

For the year 1981, Strathmore had 98% boiler system availability, and 97% overall system availability inclusive of FGD system.

Pilot-plant research

The dry FGD expertise upon which Koch Engineering based the successful Strathmore improvements derives largely from extensive pilot-plant research conducted at Abcor. The Abcor pilot facility rounds out Koch's three different scales of dry FGD research: the full-scale industrial-sized system at Strathmore [dryer volume of 125 m^3 (4,418 ft^3)]; the semiworks test system at Wichita [dryer volume of 61 m^3 (2,138 ft^3)]; and the pilot unit at Abcor [dryer volume of 3.1 m^3 (109.1 ft^3)].

The two smaller systems (Abcor and Wichita) have been used to develop a scale-up methodology, by which full-size industrial systems could be designed completely by scaling up pilot-plant results. This methodology has been corroborated by applying it to a scale-up of Abcor pilot-plant data to Strathmore, and then taking measurements at Strathmore to verify agreement. Figure 8, which compares Abcor results to Strathmore results, indicates excellent agreement and validates the scale-up methodology.

In addition to scale-up research, the purposes of the Abcor dry FGD pilot facility are:

1. To develop, improve, and optimize the dry FGD system.
2. To simulate actual on-line boiler conditions including continuous run, startup, shutdown, and sudden flow and operating changes; and to examine their influence on system operation.

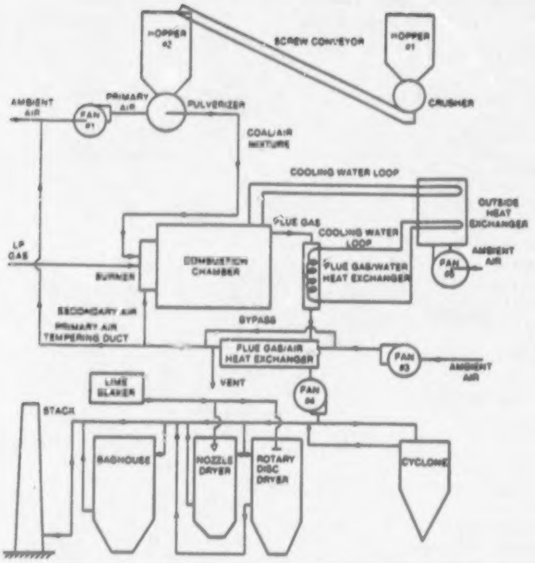


Figure 9. Abcor pilot plant.

3. To execute tests for clients with their coal, lime, and operating conditions, thus verifying the ability to achieve required SO_2 removal.

4. To develop superior atomization and gas/liquid mixing devices.

5. To study and quantify the effects of fly ash, reheat and product recycle.

6. To compare, first hand, centrifugal discs with two-fluid nozzles.

The pilot plant consists of: a 1.5 MW (5 million Btu/h) pulverized coal combustion system (complete with coal-storage facilities, coal crusher, coal pulverizer, burner, combustion chamber, heat exchangers, fans, and pumps); a two-fluid nozzle spray-dryer; a rotary disc spray dryer; a cyclone for particulate removal; a pulse-jet baghouse; a lime slaking facility; and conjunctive equipment and monitoring instrumentation, Figure 9.

The *nozzle spray dryer* is 1 m (40 in.) I.D. by 4.3 m (14 ft 2 in.) (straight wall) high, with a 0.8 m (30 in.) high bottom cone. The volume is 3.1 m^3 (109.1 ft^3). A single two-fluid nozzle is centered at the top. Either air or steam may be used as the atomizing fluid. View ports and hatches on the tower permit observation of the spray and mixing. A proprietary gas/spray mixing device is located within the dryer. The entire tower is insulated and is instrumented with thermocouples and pressure taps. The tower was designed and built by Koch Engineering.

The disc spray dryer is 1.5 m (60 in.) I.D. by 3.1 m (10 ft) (straight wall) high, with a 0.8 m (30 in.) high bottom cone. The volume is 6 m^3 (212.7 ft^3). A single rotary disc atomizer is centered at the top. The dryer is insulated and is also instrumented with thermocouples and pressure taps. Either the nozzle or the rotary disc dryer may be used for testing SO_2 removal efficiency, through the use of bypass ducts and dampers. The disc dryer was designed and built by Koch Engineering.

The *baghouse* is a pulse-jet type collector, with 42 bags and a filter surface area of 36.7 m^2 (395 ft^2). The pressure drop across the baghouse is typically 0.5–1 kPa (2–4 in. w.c.). Reverse air-cleaning pulses are controlled by an adjustable timer.

Currently, *lime is slaked* batchwise by mixing a specified weight of quicklime (CaO) with tap water. The slaked lime is fed to the nozzle by a progressive cavity pump. Shortly, this batch-type slaking procedure will be replaced with a continuous paste-type slaking process, in which lime and water will be metered continuously at a controlled rate. The lime will be fed by a screw feeder to an Abcor-designed paste slaker. The slaked lime will be degritted by a vibrating screen, and the degritted lime will be pumped to the atomizer.

Located within the lime slaking room is an *abrasion test loop.* In this system, several test nozzles are continuously monitored for energy consumption, abrasion, and plugging while spraying lime slurry and flyash into a recirculation tank. The test loop is useful in the development of new abrasion-resistant, nonplugging, energy-efficient nozzles. The system is also used for testing performance of slurry pumps and conjunctive hardware.

SO_2 concentration is measured at the dryer inlet, at the dryer outlet, and at the baghouse outlet. The monitoring instrument is an UV photometric SO_2 monitor, which is also used at Strathmore.

The system has been automated with a *microprocessor controller,* which controls dryer outlet temperature, atomizing air pressure, percent O_2 in flue gas, and lime slurry concentration according to operator set-points. It also provides continuous, on-line data-acquisition, self-checks, and warnings when monitored parameters go out of range.

A critical component of a *nozzle-based* dry FGD system is the nozzle, which must atomize effectively (to obtain 20–50 μm droplets) at low energy consumption. It must also be abrasion-resistant, nonplugging, easy to clean, and inexpensive. Since such a nozzle did not appear to be commercially available, Koch Engineering and Abcor have developed a proprietary nozzle that fulfills these requirements. Current tests indicate that the Abcor nozzle may be expected to last for more than 2,000 hours of continuous operation with no sign of abrasion or plugging.

In August of 1982, the external mix nozzles at Strathmore were replaced with Abcor internal mix nozzles. These nozzles have a liquid capacity of 0.3 L/s (4.5 gpm) each, and a measured air consumption of 0.17 m^3/L (22.5 standard ft^3/gal) [at 551 kPa (80 psig)] of slurry atomized (corresponding to approximately 1 HP per 1000 ACFM of flue gas). This reduced atomization power requirements by 65%. The nozzles have shown no signs of abrasion or plugging, and represent power costs comparable to those of rotary discs.

Another critical component is the flue gas/spray droplet *mixing device*. This device must insure intimate contact between the gas and the spray, and it must do so at low pressure drop. In addition, it must yield a flow pattern that matches the tower geometry, to prevent wetout. Koch Engineering and Abcor have developed a proprietary mixing device that meets all these criteria, permitting operation of the spray dryer within −12°C (10°F) of adiabatic saturation at a pressure drop of less than 250 Pa (1 in. w.c.).

Experiments

In the Abcor pilot plant, we have studied the influence of the following parameters on the system SO_2 removal efficiency: stoichiometric ratio ϕ (moles of absorbent per mole of SO_2 at the dryer inlet); approach to adiabatic saturation ΔT_{a-s}, sprayed droplet size d; residence time within the dryer τ; absorbent (calcium or sodium-based); inlet SO_2 concentration; flue gas temperature; and lime slaking parameters.

Of these, the most influential are stoichiometric ratio ϕ and approach to adiabatic saturation ΔT_{a-s}. It is important to note, however, that atomization, gas/liquid mixing, and residence time are all coupled and critically important in obtaining optimum system performance.

In the system, ϕ and ΔT_{a-s} are changed by varying the concentration of absorbent in the absorbent mixture and by adjusting the L/G, or gallons of absorbent mixture sprayed in per thousand standard cubic feet of flue gas. Dryer residence time τ is varied by adjusting the flue gas flow rate G, and sprayed droplet size d is adjusted by varying the nozzle atomizing air pressure, p_{air}.

The stoichiometric ratio ϕ is calculated from the following equation:

$$\phi = (6{,}908)\,(L/G)\,(CaO)/(SO_2)$$

where

ϕ = stoichiometric ratio, mol of CaO per mol of SO_2
L = slurry flow rate, (gal/min)
G = flue gas flow rate, thousand standard (ft^3)/min
(SO_2) = flue gas SO_2 concentration at dryer inlet, ppm
(CaO) = slurry concentration, (lb) of CaO per (gal) of slurry

The liquid flow rate L is measured by a coriolus force mass flowmeter, which is calibrated monthly. The gas flow rate G is measured by annubars on primary and secondary air lines and is checked with a 152 mm (6 in.) diameter square-edged calibrated orifice at the baghouse outlet. The SO_2 concentration is measured with an UV photometric SO_2 monitor, which is calibrated twice daily.

The slurry concentration is measured by double titration of the slurry with 1.0 N HCl to a phenolphthalein endpoint. The approach to adiabatic saturation ΔT_{a-s} is determined by subtracting the wet-bulb temperature T_s from the dryer outlet temperature T_o. Flue gas oxygen concentration is measured with an O_2 analyzer. Atomizing air pressure is measured with conventional regulators and pressure gauges. Pressures throughout the system are measured with magnehelic gauges, ΔP's are measured with inclined manometers, and temperatures with thermocouples and digital readouts.

Figure 8 presents the results of pilot-plant experiments conducted at an approach to adiabatic saturation of about −17°C (30°F) and at varying stoichiometric ratios (solid line). Superimposed on this plot are data obtained at Strathmore under similar conditions (open circles). The excellent agreement is an indication of the accuracy and precision of the data, and substantiates the scale-up criteria that have been developed for the design of full-scale systems from pilot-plant data.

Thus, the Abcor dry FGD pilot plant has already demonstrated a very significant utility. It has enabled the development of scale-up criteria, as well as of superior two-fluid nozzles and mixing devices. It has verified the performance of a dry FGD system in conjunction with an operating coal-fired boiler. It has also provided the background data and expertise that were essential for solving the problems at Strathmore and for making that the first full-scale operating dry FGD system in North America.

In conclusion

The combination of a pulverized coal-fired industrial boiler and a dry SO_2 scrubber is a cost-effective steam-generating alternative in capacities above 23 Mg (50,000 lb) of steam per hour. This combination represents the most responsive and flexible way to utilize noncompliance coal over fast-changing load conditions. Its operating simplicity is the yardstick by which all other coal-burning technologies must be measured.

Although dry FGD technology has only recently been developed, the commercial feasibility of dry SO_2 scrubbing technology has been demonstrated at the Strathmore Paper Co., the first full-scale operating dry flue gas desulfurization system in North America. There a commercial dry flue gas desulfurization system has passed all EPA (Environmental Protection Agency) compliance tests; it has been operating since July, 1979 with more than 20,000 hours on-line and an availability of more than 97%. ■

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27a

J. P. Menat is manager of the Environmental Engineering Dept. at Abcor, Inc. The author of more than 10 articles on combustion, nitric oxide formation, laser diagnostic techniques, air pollution control, industrial hygiene monitoring, domestic energy conservation, and flue gas desulfurization, his current interests include solar energy, dry flue gas desulfurization, mist elimination from gas streams, and domestic/industrial energy conservation. He earned his B.S.E. degree in aerospace and mechanical sciences at Princeton Univ. and his M.S. and Ph.D. degrees in environmental engineering at Stanford Univ.

28a

M. Mutsakis is division manager of Koch Engineering's Air Pollution Control Group. The holder of a B.S.Ch.E. degree from Pratt Institute, a M.S.Ch.E. degree from M.I.T. and an M.B.A. in management from New York Univ., he has authored numerous articles on air pollution control systems and static mixing equipment.

29a

This Electric Power Research Institute article describes, in general terms, nitrogen oxides as pollutants and opportunities for their control. Special emphasis is placed on burner and furnace design, which can substantially reduce emissions. Other processes to control NOx, such as flue gas treatment, are also discussed.

Controlling Oxides of Nitrogen

by J. Edward Cichanowicz

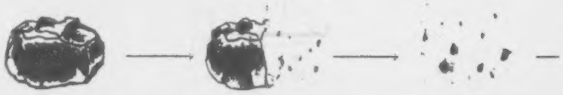
Nitrogen's traditional fate is giving way to a controlled destiny through advances in coal combustion chemistry, burner and furnace design, and flue gas treatment.

Two categories of emissions generated from the combustion of coal have come under scrutiny, which has led to the development of control technologies. Electrostatic precipitators are well established in dealing with particulate matter, and scrubbers are coming into wider use for controlling sulfur dioxide. New developments are improving the performance and reliability of both. As emissions of these two pollutants are reduced to required levels, it appears that more regulatory attention will be directed to controlling oxides of nitrogen (NO_x).

Emission standards

The health effects and public welfare issues of NO_x emissions aren't yet clear. Nevertheless, while they are being defined, New Source Performance Standards (NSPS, which set limits on pollutant emissions from new power plants) are becoming more stringent. The requirement has been that no more than 0.7 lb of NO_x be emitted for every million Btu of heat released; the new maximum issued late in May is 0.6 lb or 0.5 lb, depending on coal properties. The latter figure represents a significant (about 60%) reduction from uncontrolled levels.

Furthermore, NSPS are expected to be tightened in future years. An Environmental Protection Agency (EPA) report concludes that reduction to 0.2 lb/10^6 Btu would be required to maintain nationwide mass emissions at current levels. But even NSPS are not the sole NO_x emission criteria. Still lower values may now be specified if the ambient nitrogen dioxide (NO_2) concentration exceeds local air quality standards. Also, federal legislation is being considered for certain unspoiled areas that would permit essentially no deterioration of their present air quality—that is, today's concentrations of NO_2 and other defined pollutants. Concern over other phenomena, such as secondary particulate formation and acid rain, may further influence NO_x standards. Taken together, these factors could lead to regulatory definition of control technology beyond the capability of any that has been demonstrated to date. The potential for NO_x control and its attendant economic implications are therefore under urgent investigation.



Raw coal at the burner contains many complex nitrogen compounds . . .

that volatilize as the coal temperature rises steeply toward its combustion point . . .

yielding NO, HCN, NH_3, and other gases, plus nitrogen-bearing fuel fragments.

Formation of NO_x

NO_x refers to the sum of nitric oxide (NO) and NO_2. Coal combustion produces these from two sources: molecular nitrogen in the combustion air, and complex nitrogen compounds in the fuel itself. The NO and NO_2 that evolve from nitrogen in the combustion air are called thermal NO_x because the net rate of their formation depends strongly on combustion temperature. Oxides evolved from fuel nitrogen are called fuel NO_x. Their formation is influenced most by the chemical environment of the combustion reaction zone.

In addition to this flame-zone environment, the formation of fuel NO_x depends on fuel properties. Most coals contain relatively significant quantities of nitrogen, usually between 1.0 and 1.4 weight percent of the fuel. Typically, as a result, between 60% and 80% of all NO_x generated from the combustion of coal is attributable to fuel nitrogen. Although the task of reducing NO_x emissions requires attention to both thermal and fuel sources, most of the effort is directed at controlling fuel NO_x.

Combustion practice

For firing in a utility boiler, coal is first pulverized to a size range of about 0.003–0.05 in, then injected into the furnace

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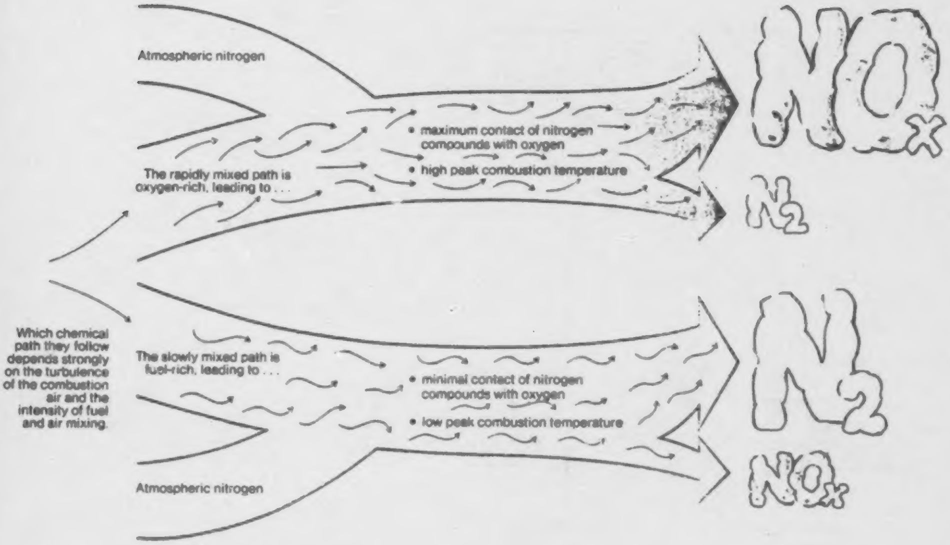

Atmospheric nitrogen
The rapidly mixed path is oxygen-rich, leading to . . .
• maximum contact of nitrogen compounds with oxygen
• high peak combustion temperature
NOx
N2
Which chemical path they follow depends strongly on the turbulence of the combustion air and the intensity of fuel and air mixing.
The slowly mixed path is fuel-rich, leading to . . .
• minimal contact of nitrogen compounds with oxygen
• low peak combustion temperature
N2
NOx
Atmospheric nitrogen

by a high-velocity air stream. The pulverized coal is injected at several locations, along with the remainder of the air required for combustion. The coal is subsequently burned in what is referred to as a *turbulent diffusion flame,* which governs the mixing and burning of coal and air; thermal energy radiates to the walls of the furnace. Combustion products, still at about 1090°C (2000°F), then exit the furnace and pass through a series of heat exchangers, where additional energy is extracted by convection.

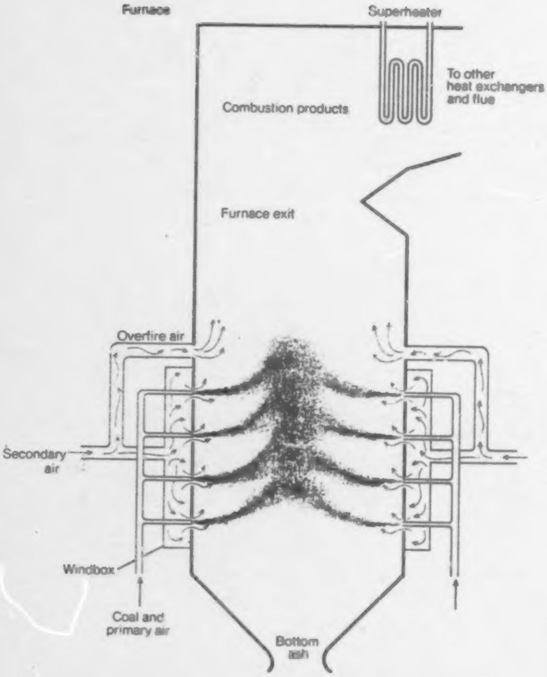
Before NO_x emissions became a factor, burner and furnace design called for the coal and combustion air to react as completely and quickly as possible—so long as temperatures were low enough to avoid slagging or fouling by coal ash. This practice provided the most efficient and economical steam generation.

Emergence of NO_x controls

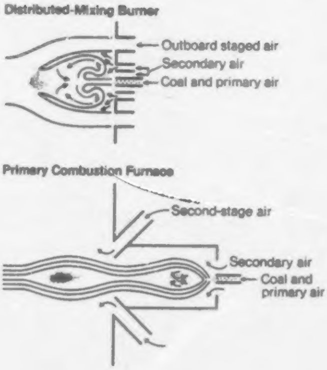
Burner design criteria are now quite different. Because it is known that rapid fuel-air mixing and high peak flame temperatures produce high NO_x levels, modern practice dictates a slowly mixed, less intense, and thus cooler flame.

Controlling thermal NO_x is mainly a matter of manipulating combustion conditions to suppress flame temperature. This is achieved with burner designs that delay the mixing of coal particles in the combustion air. Since complete combustion takes longer, more thermal energy is lost to the furnace walls from the flame zone and the ultimate combustion temperature is lower.

The control of fuel NO_x also relies on delayed mixing. The ultimate fate of nitrogen-containing fuel fragments—to be oxidized to unwanted NO_x or to be reduced to molecular nitrogen (N_2)—depends on the abundance or deficiency of oxygen encountered in the combustion zone, where the parent fuel particles


Furnace
Superheater
To other heat exchangers and flue
Combustion products
Furnace exit
Overfire air
Secondary air
Windbox
Coal and primary air
Bottom ash

Today's opposed-firing utility furnace may have four tiers of burners, with turbulent diffusion of coal and air induced by the flow of secondary air through an annulus surrounding each burner tip. The relatively smooth flow pattern delays fuel-air mixing and extends the time for combustion so that the peak temperature is reduced. This necessitates a somewhat larger furnace wall area for thorough heat transfer, but it also suppresses NO_x formation. Overfire air, injected closer to the furnace exit, further encourages fuel-rich combustion and controls peak temperature. Carbon burnout is essentially complete as gases leave the furnace.


Distributed-Mixing Burner
Outboard staged air
Secondary air
Coal and primary air
Primary Combustion Furnace
Second-stage air
Secondary air
Coal and primary air

Delaying, or staging, coal combustion to suppress NO_x formation is achieved quite differently in these two advanced burner designs now under development. Fuel and air mixing patterns and the general shape of the heat-release zones are shown as they are presently thought to occur.

In the Distributed-Mixing Burner, primary air injects the coal into the flame zone, where secondary air progressively contacts the eddying fuel particles. Outboard staged air leads to complete burnout and envelops the flame zone to protect furnace surfaces from interim combustion products that are corrosive.

The Primary Combustion Furnace physically confines the initial, fuel-rich flame zone of each burner to a chamber outside the main furnace. Two distinct heat-release zones result, and special lining materials for the primary furnace resist corrosion attack there.

first break down to produce a variety of intermediate species. If the environment is fuel-rich (oxygen-deficient), these species are mainly hydrogen cyanide (HCN), ammonia (NH_3), and NO_x. If not overly exposed to oxygen, these will subsequently reduce to N_2 and other byproducts. However, if sufficient oxygen is encountered, they will oxidize to (or remain as) NO_x.

Minimizing fuel NO_x emissions thus requires minimizing the amount of fuel and fuel-rich intermediate species that rapidly mix with oxygen. This is achieved by burner design schemes that delay the contact of fuel and air. Mixing can be even further retarded by modifying furnace design to permit injecting a portion of the combustion air (about 20%) above the burners, closer to the furnace exit. Injecting this "overfire air" also tends to reduce NO_x emission levels.

The state of the art

How effective have these design refinements been in reducing NO_x emissions? Uncontrolled NO_x emissions were once as high as 1.0–1.3 lb/10^6 Btu. Now, in units designed for NSPS compliance, emissions are in the range 0.5–0.7 lb/10^6 Btu, depending on furnace design and coal properties.

What have been the penalties? Increased burner complexity is reflected in higher capital costs for steam generators and in operating and maintenance requirements. However, costs attributable to NO_x controls have been a relatively small percentage of steam generation costs. NO_x reduction measures to date can therefore be judged as relatively cost-effective.

Can the technique of delayed fuel-air mixing be used for even greater NO_x suppression? Not without the development of radically different hardware, it appears. Fuel utilization becomes poor as combustion temperatures are lowered beyond those required for adequate fuel burnout. In addition, a new problem arises: the potential for corrosion of furnace walls due to exposure to combustion products.

Advanced control concepts

Two current equipment developments provide the potential for emissions as low as 0.2 lb/10^6 Btu, without efficiency penalties or corrosion problems. These innovations rely on the same physical and chemical principles as commercial designs. One option is called the Distributed-Mixing Burner, under investigation by Energy and Environmental Research Corp. (EERC). This option has federal sponsorship through EPA. The other is the Primary Combustion Furnace, under development by Babcock & Wilcox (B&W), and cosponsored by B&W and EPRI.

In EERC's design the coal is initially fired in an extremely fuel-rich mixture—essentially, no free oxygen. Very high turbulence is created by the burner configuration, so that incoming coal particles are rapidly and thoroughly mixed with partially burned nitrogen-bearing fuel fragments. This environment favors chemical reduction of nitrogen compounds to molecular N_2. Combustion is completed only as gases leave the fuel-rich zone and mix relatively slowly with the remainder of the combustion air that is injected through ports distributed around the burner. Air flow patterns are designed to provide a protective envelope that isolates the corrosive fuel-rich zone from the furnace walls.

Tests at the pilot scale under well-controlled laboratory conditions are encouraging; EPA reports potential NO_x levels of less than 0.2 lb/10^6 Btu. Conclusive results await full-scale tests in an operating utility boiler. These tests are currently being planned and are expected to begin in 1980.

B&W's Primary Combustion Furnace is a parallel effort that employs a physically separate first stage to produce the fuel-rich conditions essential for reducing the nitrogen-containing fuel fragments to N_2. Only at the entrance to the second stage is more air added, thereby permitting complete carbon burnout while maintaining the physical and chemical conditions for minimal NO_x formation. Corrosion potential in this furnace configuration is managed by confining the fuel-rich zone to the first stage, where corrosion-resistant materials can be used without adding significantly to overall furnace cost.

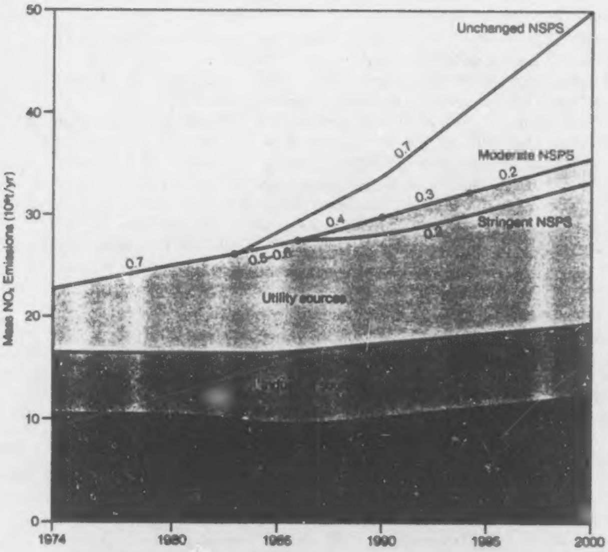
Pilot-scale results from B&W's Research Center are comparable to those for EERC's Distributed-Mixing Burner. In this case also, greater certainty awaits the results of scale-up of the Primary Combustion Furnace, first to a larger pilot unit (now being tested) and then to a utility boiler demonstration. Both approaches are likely to be more expensive than today's combustion NO_x controls. Assessment of the utility-scale demonstrations will define the overall economics and allow the cost-effectiveness of these control concepts to be determined.

NO_x removal from flue gas

Will it become necessary to lower NO_x emission levels below 0.2 lb/10^6 Btu? The rationale for more stringent emissions standards has been noted. In fact, NO_x control equivalent to levels as low as about 0.10 lb/10^6 Btu has been promulgated for California's South Coast Air Basin, to be implemented in the early 1980s. From what is known about combustion control approaches today, it appears unlikely that they alone are feasible for the purpose. In response to more stringent standards, postcombustion flue gas treatment (FGT) has come under study.

More than 50 FGT processes for NO_x control have been identified. "Dry" processes employ a gaseous reagent that selectively reduces flue gas NO_x to molecular N_2. "Wet" processes, on the other hand, employ an aqueous reagent that first absorbs and then reduces NO_x for disposal in liquid form. Several generalizations can be made about the relative merits of the two kinds of processes.

Dry processes tend to require less complex equipment, partly because there is no liquid waste stream for disposal. Costs therefore appear to be lower, for both initial capital requirements and subsequent operation and maintenance. And



The total tonnage of NO_x emitted by three major U.S. categories of combustion will be heavily influenced by growth in utility generation, the fuels chosen to achieve it, and the unit emission standards applied. These curves illustrate futures in which utility steam-electric generation rises 6% annually, with three-fourths of the new and replacement capacity fueled by coal (the rest by nuclear energy).

Evolution of New Source Performance Standards is assumed, as indicated by the unit values for NO_x emission (lb/10^6 Btu) shown with each curve. The moderate NSPS curve reflects the NO_x standard issued in May 1979 and its possible effect on mass emissions between 1983 and 1986.

denitrification efficiency (the percentage reduction of NO_x entering the process) appears to be superior, because of the relative simplicity of uniformly dispersing a gaseous reducing agent in the flue gas. Dry FGT processes have therefore had the most R&D attention, initially in Japan and mainly for dealing with NO_x from low-sulfur fuel oil and natural gas, which are widely used in that country. The major problem is to adapt these processes to coal-fired applications, where higher levels of sulfur dioxide (SO_2) and particulate matter can easily impair their performance. (Because these processes require a temperature range of 290–430°C [550–800°F] for optimal performance, FGT equipment for NO_x removal must be located upstream from air preheaters, precipitators, and SO_2 scrubbers in the exhaust gas duct.)

Selective catalytic reduction

Of most interest and potential commercial feasibility are several dry-process FGT approaches that use NH_3 as the reducing agent. These processes employ catalysts to speed the reactions, so that selective NO_x reduction can occur throughout the range of flue temperatures encountered under practical operating conditions, and are termed selective catalytic reduction (SCR) processes. These processes differ in catalyst composition and physical characteristics, but most developers claim denitrification efficiencies of 90%. Exact figures depend on catalyst design, process temperature, the ratio of NH_3 to NO_x, and other factors in boiler operation. Although these findings encourage the view that SCR is technically feasible, there are unresolved problems that could limit its practical performance and degrade its cost-effectiveness, which, even at optimistic levels, is poor.

Excess NH_3 carryover appears to be the foremost problem; NH_3 that does not participate in the reduction process is emitted, possibly constituting a separate environmental hazard. Sulfur trioxide (SO_3) is another problem; it is easily formed from SO_2, which is at high levels in the flue duct where the FGT equipment is located. This, too, can have environmental implications. Furthermore, SO_3 can lead to severe corrosion—by itself, or through combination with water to make an acid mist, or in reactions with NH_3 to produce several corrosive compounds (ammonium sulfate and ammonium bisulfate). Air preheaters, for example, would suffer from corrosion or from plugging by these compounds.

NH_3 demand is an important consideration in assessing the feasibility of SCR processes. For a single 500-MW plant, assuming 90% NO_x reduction from an initial level of 0.6 lb/10^6 Btu by an SCR process, the annual NH_3 requirement could be almost 6000 tons. The production of NH_3 requires a methane supply on a one-for-one molecular basis. And since methane must be derived from either natural gas or coal gasification, the energy and economic implications of NH_3 requirements can't be ignored. Catalyst life is another factor. For most SCR processes, it is about one year; so this raises additional problems of supply and disposal—and their costs.

In sum, FGT for NO_x control is characterized by many unanswered questions and unresolved issues. System costs are therefore hard to assess. Current estimates place them in the range of $75–$200 per kilowatt of plant capacity, several times the cost of even advanced combustion NO_x controls and thereby a significant portion of total plant costs.

FGT at pilot scale

The reality is that FGT may be required in regions that mandate stringent NO_x emissions standards. Decisions whether to promulgate such standards can best be made in the light of accurate and detailed assessments of the relative benefits and burdens of FGT systems. To provide such information for use in the decision-making process, three studies will soon be under way that will carefully examine all aspects of FGT technology.

The largest pilot facility to be tested, and for the longest evaluation period, will involve a commercial dry SCR process developed jointly by Kawasaki Heavy Industries and Japan's Electric Power Development Corp. Sponsored by EPRI, these tests will be initiated in January of 1980 at EPRI's Advanced Emissions Control and Test Facility at Denver, Colorado. The facility draws flue gas from the adjacent coal-fired Arapahoe station operated by Public Service Co. of Colorado. In this test a gas flow equivalent to that from 2.5 MW of generating capacity will be treated in studies of denitrification efficiency under authentic plant conditions. Extending over a year, the test program will provide detailed measurements of NH_3 emissions, SO_3 concentrations, corrosive compounds, catalyst degradation, and other factors in an attempt to define total system impact and costs.

Concurrently, EPA will sponsor field tests of two commercially available SCR processes. The Hitachi-Zosen process is to be installed at Georgia Power Co.'s Plant Mitchell; and the Universal Oil Products process will be installed at Tampa Electric Co.'s Big Bend station. Both will be used to treat flue gas streams equivalent to those from about 1 MW of capacity during a 6–9-month study that should begin late this year.

Toward conclusive knowledge

The conventional and prospective NO_x control technologies discussed here differ in their ability to restrict power plant NO_x emissions and in their foreseeable cost-effectiveness. With today's level of understanding, it appears that the greatest promise for effective NO_x control lies in preventing NO_x formation in the furnace. If further reductions that are beyond the reach of this strategy alone are necessary in some air quality regions, FGT will also be needed.

Efforts that are now under way to define the environmental benefits and cost impacts of NO_x controls will provide the necessary information in the next several years to enable utilities and national and local legislative bodies to work toward the best solutions regarding NO_x emissions and public health and welfare.

This article in Hydrocarbon Processing is a review of the state-of-the art processes available for NOx controls. Especially interesting are Tables 7 and 8, (p. 378), comparing facilities in the U.S. and Japan. The Japanese have refined the control technologies, many of which were originally developed in the U.S.

NO_x controls in review

Advances in NO_x control have been extensive since the last look. Much work has been done to improve the situation.

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MODIFYING THE COMBUSTION PROCESS is becoming the most widely used technology for reducing combustion generated oxides of nitrogen (NO_x). The current trend in the United States is to apply combustion controls first in meeting NO_x emission standards. However, the Japanese have developed and commercialized a number of Flue Gas Treatment (FGT) systems because of Japan's more stringent NO_x standards.

NO_x Formation. Oxides of nitrogen (NO_x) formed in combustion processes are usually due either to thermal fixation of atmospheric nitrogen in the combustion air, leading to "thermal NO_x," or to the conversion of chemically bound nitrogen in the fuel, leading to "fuel NO_x." For natural gas firing, nearly all NO_x emissions result from thermal fixation. With residual oil, crude oil, and coal, the contribution from fuel-bound nitrogen can be significant and under certain operating conditions, predominant.

Thermal NO_x. Fixation of a small fraction of the molecular nitrogen in combustion air results in formation of thermal NO_x. The rate of formation is strongly temperature dependent, generally accepted to be in excess of 3,200°F in the presence of excess air.

At high temperature, the usually stable oxygen molecule, O_2, dissociates to oxygen atoms which are very reactive. Oxygen atom, O, being very reactive attacks the otherwise stable nitrogen molecule, N_2, to form NO_x.

The most widely accepted reactions that describe the formation of thermal NO_x are:

$$O_2 + M \rightleftarrows O + O + M$$

$$N_2 + O \rightleftarrows NO + N$$

$$N + OH \rightleftarrows NO + H$$

$$O_2 + N \rightleftarrows NO + O$$

Fuel NO_x. The role of fuel-bound nitrogen as a source of NO_x emissions from combustion sources has been recognized recently.[2] Although the relative contribution of fuel and thermal NO_x to total NO_x emissions from sources firing nitrogen-containing fuels has not been well established, recent studies[3,4] indicate that fuel NO_x is significant and may even predominate. One recent study[4] showed that fuel NO_x can account for over

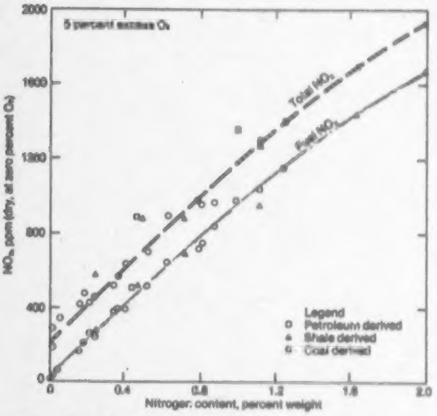


Fig. 1 — NO_x emissions from liquid fuels.[5]

TABLE 1 — NO_x Control Methods

Reduction of NO_x Formation	NO_x Removal From Flue Gas
• Combustion Modification	• Dry Process
Low Excess Air Combustion	Selective Catalytic Reduction (SCR)
Flue Gas Recirculation	Selective Noncatalytic Reduction (SNR)
Staged Combustion	
Low NO_x Burners	
Catalytic Combustion	
• Use of Low Nitrogen Fuel	• Wet Process
Change of Fuel	Oxidation Absorption
Nitrogen Removal of Fuel	Oxidation Reduction
	Absorption Reduction

TABLE 2 — New source performance standards for steam generators

Fuel	lb. NO_x/MMBTU
Coal	
Bituminous	0.6
Subbituminous	0.5
Lignite	
Pulverized Fired	0.6
Cyclone Fired	0.8
Coal Derived Liquids	0.5
Oil	0.3
Gas	0.2

50 percent of total NO_x production from residual oil firing and approximately 80 percent of total NO_x from coal firing. Effect of fuel nitrogen content on NO_x emissions is illustrated in Fig. 1.

EMISSION CONTROL

NO_x control techniques for stationary sources operate either through suppression of NO_x formation or through physical or chemical removal of NO_x from stack gases. Major techniques for NO_x emission control are listed in Table 1.

NO_x formation can be reduced by:

- Reducing nitrogen level at peak temperature
- Decreasing oxygen level at peak temperature
- Reducing peak temperature and residence time in the combustion zone.

This can be accomplished by employing several combustion modification techniques such as those listed in Table 1.

Combustion Modification. This, the most cost-effective and energy efficient technology used to control combustion-generated oxides of nitrogen, has been successfully implemented on existing gas and oil fired boilers to comply with emission standards (Table 2). Results show up to 60 percent reduction in NO_x emissions.

Combustion modification techiques can be classified into three major categories:

- ► Change operating conditions
  - Low-excess air combustion
  - Change fuel air contact in combustion chamber
  - Reduce heat load in combustion chamber
  - Lower air preheat temperature.
- ► Modification burner design
- ► Modification combustion system design
  - Reduce heat in combustion chamber (increase furnace volume)
  - Staged combustion
  - Flue gas recirculation
  - Water or steam injection.

Low Excess Air Combustion. Reducing the total amount of excess air supplied for combustion in industrial boilers and process heaters decreases both thermal and fuel NO_x formation. Since low excess air firing also increases efficiency (decreased fuel consumption) it is generally considered as part of an energy conservation program and can easily be implemented. The ultimate level of excess air is limited by smoke or carbon monoxide emissions in the stack. NO_x reductions of 17 to 38 percent can be achieved by reducing excess air to the minimum. Low excess air firing (below 5 percent) is becoming standard practice on most oil and gas boilers.

Flue Gas Recirculation. Recirculation of flue gas to the primary combustion zone reduces thermal NO_x formation by lowering peak flame temperature and excess oxygen level. A typical effect of the flue gas recirculation of NO_x emission is shown in Fig. 2. Since this method mainly reduces thermal NO_x, its application will, therefore, be restricted to low nitrogen fuels, such as natural gas, distillate oil, and low nitrogen residual oils.

This scheme is usually not cost-effective due to the increased energy costs due to reheating the recirculated flue gas.

Staged Combustion is based on operation with a rich primary combustion zone in the furnance to reduce

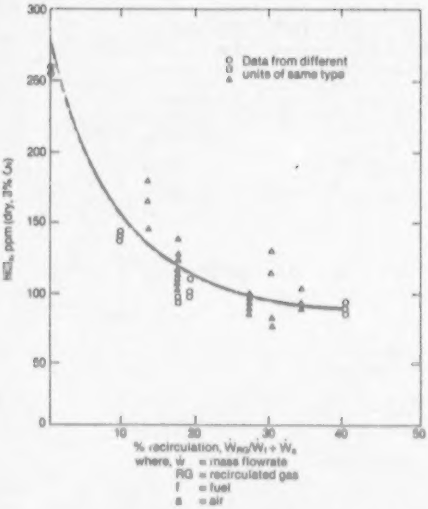


Fig. 2 — Effect of flue gas recirculation on NO_x emissions.[6]

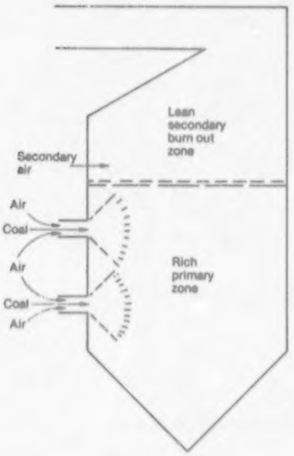

Lean secondary burn out zone
Secondary air
Air
Coal
Air
Coal
Air
Rich primary zone

Fig. 3 — Staged combustion.

oxygen availability and peak temperature, followed by secondary air injection to achieve carbon burnout. Reduced oxygen availability lowers conversion of fuel nitrogen to NO and reduced peak temperature with subsequent heat removal prior to secondary air addition, reduces thermal NO_x (Fig. 3). Air supplied to the burners is less than that necessary to completely burn the fuel. This produces a fuel-rich primary zone. Secondary air is added above the top row of burners and an equal or greater secondary zone residence time is provided for carbon burnout. A reduction of 30 to 50 percent in NO_x emissions can be achieved in existing industrial boilers.

Low NO_x Burner. Peak flame temperatures, and residence time in the fuel-rich zones can be controlled by burner design parameters, such as fuel distribution and rate of air mixing. Low NO_x burner design is based upon the concept of distributed mixing. The burner is operated in a highly staged manner at about 70 percent stoichiometry. Coal and air are injected into the firebox through a burner jet and accompanied by two stages of external air (Fig. 4). A third external air stream joins this mixture after initial combustion has taken place. In this way, initial coal combustion takes place in a relatively reducing environment that keeps nitrogen oxidation to a minimum. By keeping high fuel-to-air ratio in the initial combustion, most of the fuel nitrogen is converted to nitrogen gas.

Several pulverized coal burners designed to reduce NO_x have been tested by boiler manufacturers. These studies have shown the potential for 65 to 90 percent reduction relative to uncontrolled emissions.

Low NO_x burners have also been effective in gas and oil firing applications.

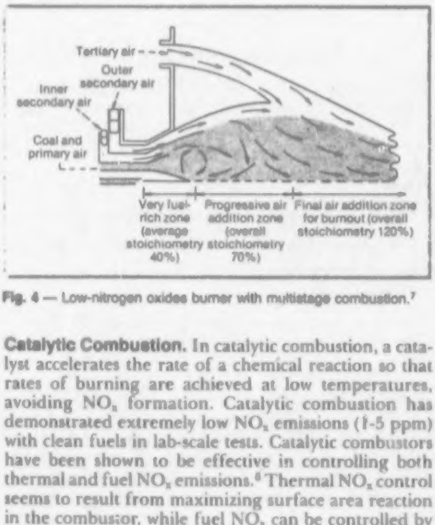

Tertiary air
Outer secondary air
Inner secondary air
Coal and primary air
Very fuel-rich zone (average stoichiometry 40%)
Progressive air addition zone (overall stoichiometry 70%)
Final air addition zone for burnout (overall stoichiometry 120%)

Fig. 4 — Low-nitrogen oxides burner with multistage combustion.[7]

Catalytic Combustion. In catalytic combustion, a catalyst accelerates the rate of a chemical reaction so that rates of burning are achieved at low temperatures, avoiding NO_x formation. Catalytic combustion has demonstrated extremely low NO_x emissions (1-5 ppm) with clean fuels in lab-scale tests. Catalytic combustors have been shown to be effective in controlling both thermal and fuel NO_x emissions.[6] Thermal NO_x control seems to result from maximizing surface area reaction in the combustor, while fuel NO_x can be controlled by operating at a rich fuel/air ratio which minimizes the formation of NO_x.

Other Techniques. Techniques which have been tested for reducing primary combustion zone temperature (lowers thermal NO_x production) include reduced air preheat and coolant (steam or water) injection. Since these techniques cause significant reductions in thermal

Table 3 — NO_x flue gas treating technology[4]

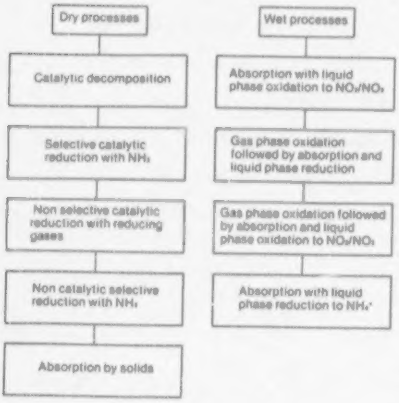
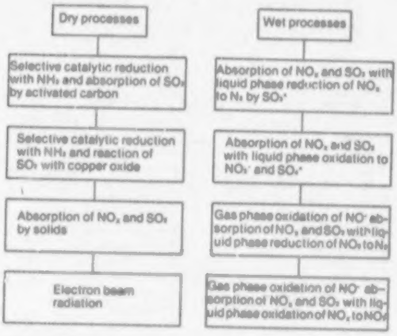

Dry processes
Catalytic decomposition
Selective catalytic reduction with NH_3
Non selective catalytic reduction with reducing gases
Non catalytic selective reduction with NH_3
Absorption by solids
Wet processes
Absorption with liquid phase oxidation to NO_2/NO_3
Gas phase oxidation followed by absorption and liquid phase reduction
Gas phase oxidation followed by absorption and liquid phase oxidation to NO_2/NO_3
Absorption with liquid phase reduction to NH_4^+

Table 4: NO_x/SO_x flue gas treating technology[4]


Dry processes
Selective catalytic reduction with NH_3 and absorption of SO_2 by activated carbon
Selective catalytic reduction with NH_3 and reaction of SO_2 with copper oxide
Absorption of NO_x and SO_2 by solids
Electron beam radiation
Wet processes
Absorption of NO_x and SO_2 with liquid phase reduction of NO_x to N_2 by SO_3^-
Absorption of NO_x and SO_2 with liquid phase oxidation to NO_3^- and SO_4^-
Gas phase oxidation of NO^- absorption of NO_x and SO_2 with liquid phase reduction of NO_2 to N_2
Gas phase oxidation of NO^- absorption of NO_x and SO_2 with liquid phase oxidation of NO_x to NO_3

efficiency their use as practical control techniques has been questionable.

Low Nitrogen Fuel. NO_x can also be controlled through fuel modifications. Natural gas firing is an attractive NO_x control strategy due to the absence of fuel NO_x in addition to the flexibility gas provides for implementation of combustion modification techniques. Despite superior cost-effectiveness of gas-fired NO_x control, economic consideration in fuel selection are dominated by the clean fuel shortage. The recent trend has been toward coal use for electric power and larger industrial processes. Fuel switching to natural gas or distillate oil does not seem to be a promising option.

Flue Gas Treating Processes. While current U.S. EPA NO_x emissions standards have essentially been met with combustion control modifications for fuel switching; more stringent regulations by states or local agencies, a new short term NO_x standard and/or limited availability of clean fuels may necessitate use of post combustion flue gas treating (FGT) processes. Since Japan has more stringent NO_x emission standards than the U.S., a major part of the development and commercial applications of NO_x FGT processes have originated in Japan.[7]

Flue gas treating processes are generally classified as wet or dry processes, with further distinctions as to NO_x removal alone or combined NO_x/SO_x removal. Dry processes usually utilize catalytic decomposition or adsorption while wet processes normally utilize absorption technology.

Since most dry FGT processes utilize catalytic decomposition or homogenous decomposition which require high temperature (600-1,800°F), they are normally integrated into the boiler or process heater ahead of the air preheater. Wet processes are usually add on in that they are installed downstream of all equipment prior to entering the stack.[9]

The major dry processes for NO_x and NO_x/SO_x removal are:[9-11]

- Selective catalytic reduction (SCR) with NH_3
- Selective (non-catalytic) reduction with NH_3
- Non selective catalytic reduction
- Adsorption

The major wet processes for NO_x and NO_x/SO_x removal are:[9-11]

- Absorption - reduction
- Absorption - oxidation
- Oxidation - absorption

All FGT process technologies currently under development are presented in Table 3 for NO_x removal processes and Table 4 for dual NO_x/SO_x removal processes. In addition, a listing of specific FGT processes offered by various vendors is presented in Table 5.

Most development has been on the dry processes, especially selective reduction with NH_3, due to the complexity and cost of wet processes. Wet NO_x removal processes are not now considered economically viable compared to dry processes.[12]

Two dry processes (electron beam radiation and absorption by alkalized alumina) are in early stages of development and will not be discussed in detail. If additional information is desired the vendors are listed in Table 5.

In general, dry processes have the following advantages over wet processes:

- Lower capital investments and lower annual revenue requirements
- Simpler processes with fewer equipment requirements
- Higher NO_x removal efficiencies (>90%)
- More extensive testing in large units (oil- or gas-fired boilers)

TABLE 5 — Flue Gas Treating Processes[19]

Dry Process	Wet Process
Selective catalytic reduction	Absorption-reduction
Asahi Glass[a]	Asahi Chemical
Eneron	Chisso Engineering
Exxon	Kureha
Hitachi, Ltd.	Mitsui Engineering and Shipbuilding
Hitachi Zosen	Pittsburgh Environmental and Energy Systems
JGC Paranox	Oxidation-absorption-reduction
Kobe Steel	Chiyoda Thoroughbred 102
Kurabo Knorca	Ishikawajima-Harima Heavy Industries
Kureha	Mitsubishi Heavy Industries
Mitsubishi Heavy Industries	Moretana Calcium
Mitsubishi Kakoki Kaisha	Moretana Sodium
Mitsubishi Petrochemical	Osaka Soda[a]
Mitsui Engineering and Shipbuilding	Shirogene[a]
Mitsui Toatsu	Absorption-oxidation
Nippon Kokan[a]	Hodogaya[a]
Sumitomo Chemical	Kobe Steel
Sumitomo Heavy Industries[b]	MON Alkali Permanganate
Takeda	Nissan Engineering
Ube	Oxidation-absorption
Unitika[b]	Kawasaki Heavy Industries
Universal Oil Products-Shell CuO[b]	Tokyo Electric-Mitsubishi HI
Nonselective catalytic reduction	Ube
The Ralph M. Parsons	
Selective noncatalytic reduction	
Exxon Thermal Denox	
Adsorption	
Foster Wheeler-Bergbau Forschung	
Radiation	
Ebara-JAERI	
Uncertain	
Princeton Chemical Research[a]	

[a] Essentially no information available.

[b] These companies have two dry processes: NO_x-only and SO_2-NO_x.

- No waste generation.

However, dry processes also have the following disadvantages:

- Greater sensitivity to inlet particulate matter levels
- Requirements for NH_3 from either an energy source (natural gas) or more expensive coal gasification methods
- Possible emissions of $(NH_4)_2SO_4$, NH_4HSO_4 and NH_3
- Higher reaction temperatures (350-400°C), which requires these systems to be placed in the power generation cycle before the boiler air preheater or auxiliary heating after the preheater.

In general, wet SO_x/NO_x removal processes have advantages as compared with the dry processes.

- Simultaneous SO_x/NO_x removal, which may provide economic advantage under certain applications
- Relative insensitivity to flue gas particulate matter
- Higher SO_x reduction (to 95%)

Major disadvantages of these wet processes include:

- Higher costs because of the insolubility of NO_2 in aqueous solutions
- Formation of nitrates and other potential water pollutants
- More extensive equipment requirements
- Formation of low-demand by-products
- Need for flue gas reheat
- Moderate NO_x removal efficiencies (about 90%)
- May be limited to flue gas with a high SO_x:NO_x ratio.

Selective Catalytic Reduction (SCR). SCR of NO_x with ammonia (NH_3) is the predominant technology used in flue gas treating processes. It has been demonstrated at 90% plus NO_x reduction in commercial applications. SCR processes are based on preferential reduction selectivity of NH_3 for NO_x. Since the presence of oxygen enhances NO_x reduction, reactions are best expressed as:

$$4\,NH_3 + 4\,NO + O_2 \xrightarrow{\text{catalyst}} 4\,N_2 + 6\,H_2O \quad (1)$$

$$4\,NH_3 + 2\,NO_2 + O_2 \xrightarrow{\text{catalyst}} 3\,N_2 + 6\,H_2O \quad (2)$$

Reaction 1 predominates since most flue gases contain 90-95% nitric oxide (NO). In commercial practice an NH_3:NO mole ratio of 1.05:1.0 to 1.1:1.0 is required to reduce NO_x emissions by 90% with a residual NH_3 concentration in the flue gas of less than 20 ppm.[11,12]

Major components for SCR processes are catalyst and reactor, an NH_3 injection system, and NH_3 storage. The temperature requirement for NO_x reduction is dependent on specific catalyst type but normally ranges from 570 to 850°F.[9,11,12] To obtain this temperature without reheat requirements, the reactor is normally located between the economizer and air preheater. When retrofitting an existing boiler or process heater this usually requires additional ducting and air blower modifications to avoid the energy cost of reheating. A flow diagram for a typical SCR process is shown in Fig. 5.

Since catalyst technology is considered proprietary by most vendors, knowledge of exact catalyst compositions is not readily available. In general, the catalyst is a base metal such as iron, vanadium, chromium, manganese, cobalt, nickel, copper or barium deposited on a carrier of alumina, titanium dioxide or silicon dioxide. While all of these catalysts are acceptable for NO_x reduction, many are damaged by SO_x in flue gas. The most stable and most widely used catalysts contain vanadium compounds on a titanium dioxide carrier.[9]

Catalyst configuration is a very important factor in SCR system design since particulates can blind catalyst sites.

For clean flue gases a packed catalyst bed has proved successful in Japan. This is usually achieved by prior SO_x scrubbing and a precipitator, but this is a very costly system.

For dirty flue gas streams (high SO_x and particulates) regenerable and moving bed processes are being developed, but parallel flow catalysts are most promising. In parallel flow catalyst beds, flue gas flows straight through open channels parallel to the catalyst surface.

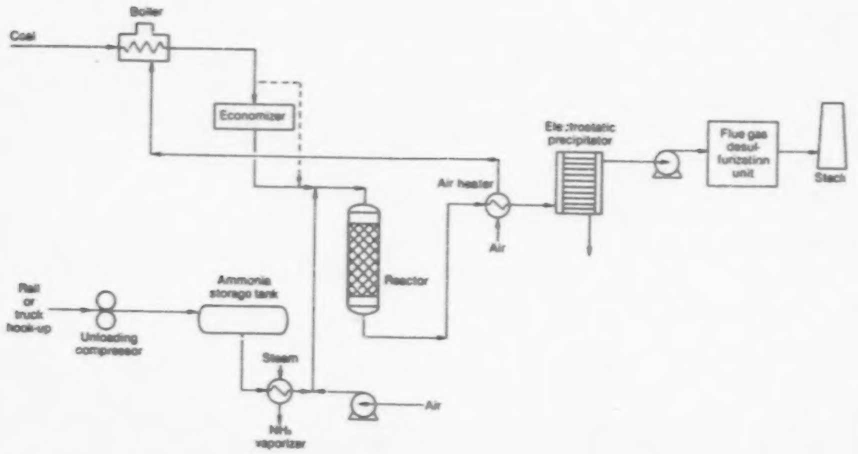

Boiler
Coal
Economizer
Electrostatic precipitator
Flue gas desulfurization unit
Stack
Air heater
Air
Reactor
Rail or truck hook-up
Unloading compressor
Ammonia storage tank
Steam
Air
NH₃ vaporizer

Fig. 5 — Typical flowsheet for NO_x selective catalytic reduction processes.[12]

Particulates in the flue gas remain entrained and avoid catalyst while NO_x reaches the catalyst surface by turbulent convection and diffusion.[12] Various parallel flow catalyst configurations are illustrated in Fig. 6.

One major concern with SCR processes is formation of ammonium sulfate and ammonium bisulfate downstream of the reactor. These compounds tend to deposit in the air preheater or other downstream equipment and cause corrosion and plugging problems. Increased soot blowing and water washing are required in most applications.

Other potential problems are excess NH_3 emissions, NH_3 availability and increased SO_3 emissions from catalytic oxidation of SO_2.

Despite these potential problems SCR processes are the most widely utilized and developed processes for NO_x removal.

Selective Noncatalytic Reduction with NH_3. Selective noncatalytic reduction (developed by Exxon) of NO_x with NH_3 is based on the same principles and chemical reactions as SCR processes, except that a catalyst is not utilized. Absence of catalyst necessitates a temperature of 1,700-1,800°F for the reactions to take place and injection of NH_3 directly into the boiler or process heater firebox.[9,13]

In addition to desired NO_x reduction reactions, there are side reactions which take place at temperatures greater than 1,800°F that produce NO, and above 2,000°F NO_x concentrations in the flue gas increase. For these reasons, NH_3 injection into the boiler or heater at the point of desired temperature is critical and can be difficult to achieve since load changes or fouling will change the firebox temperature pattern. Multiple NH_3 injection points and H_2 injection to reduce the side reactions are utilized for control.

With an NH_3:NO_x feed ratio of 1.5:1.0 to 2.0:1.0, NO_x reductions in the range of 50-60 percent have been achieved in Japan on commercial boilers. While this process is harder to control and less efficient than SCR processes, it is much less expensive to install, especially on existing boilers or heaters. A diagram of this process is presented in Fig. 7.

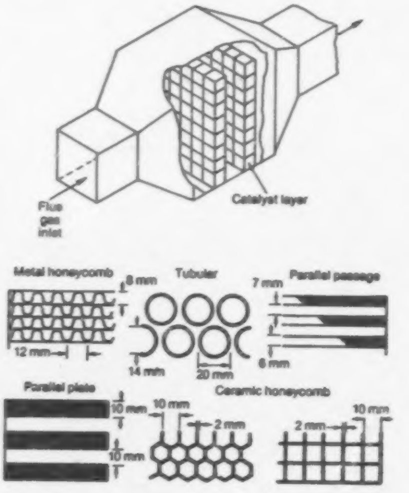

Flue gas inlet
Catalyst layer
Metal honeycomb
8 mm
12 mm
Tubular
14 mm
20 mm
Parallel passage
7 mm
6 mm
Parallel plate
10 mm
10 mm
Ceramic honeycomb
10 mm
2 mm
10 mm
2 mm

Fig. 6 — Types of parallel flow catalyst shapes for NO_x selective catalytic reduction processes.[12]

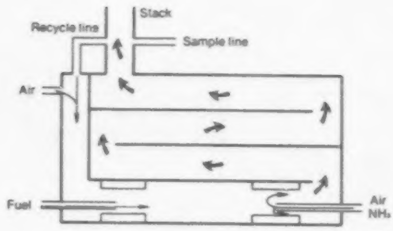


Fig. 7 — EXXON Thermal DeNO$_x$ process.[14]

Principal areas of concern are equipment corrosion and plugging due to ammonium sulfate and ammonium bisulfate byproduct formation, NH_3 emissions and NH_3 usage.

Advantages are low capital cost for installation, no catalyst blinding or plugging problems and no increased SO_3 emissions.

Shell FGT Process. The Shell Flue Gas Treating Process (licensed through UOP) is a unique combination of SO_x absorption and selective catalytic reduction of NO_x. This process utilizes a copper acceptor supported on stabilized alumina which is arranged in two or more parallel passage reactors.

The copper is first oxidized by the following reaction:[12]

$$Cu + \tfrac{1}{2} O_2 \rightarrow CuO$$

Copper oxide readily reacts with the SO_2 and O_2 in the flue gas at a temperature of 750°F to form copper sulfate:

$$SO_2 + \tfrac{1}{2} O_2 + CuO \rightarrow CuSO_4$$

Copper sulfate and copper oxide, to a lesser degree, act as catalysts for selective reduction of NO_x by NH_3. When one reactor is saturated with copper sulfate, flue gas flow is diverted to a different reactor for flue gas treatment while the saturated reactor is regenerated.

During the regeneration cycle, hydrogen is used to reduce copper sulfate to copper, as shown in the following reaction, yielding an SO_2 rich stream suitable for further processing to produce elemental sulfur or sulfuric acid.

$$CuSO_4 + 2H_2 \rightarrow Cu + SO_2 + 2H_2O$$

The reactor is then purged with steam to remove combustible gases and placed back in operation when needed.

Depending on catalyst quantity, regeneration frequency and NH_3:NO_x ratios, SO_x and NO_x reductions of 90% are possible.

Since the mechanisms for SO_x and NO_x reductions are independent, this process can be utilized for SO_x control alone by eliminating NH_3 injection, or for NO_x control alone by eliminating regeneration. A process flow diagram for this process is shown in Fig. 8.

Wet SO$_x$/NO$_x$ Processes. Since scrubbing solutions for wet NO_x removal also remove SO_x, current technology development is aimed at the dual NO_x/SO_x processes. This technology is especially well suited for high sulfur oil or coal firing since flue gas desulfurization (FGD) would be required anyway.

A major problem associated with NO_x reduction is the relatively low solubility of NO in absorbing solutions when compared to NO_2, which is much less soluble than SO_2. Since NO_x in flue gas is 90-95% NO, it is difficult to obtain high levels of NO_x removal without prior oxidation to NO_2 in either the liquid or gas phase.[11] Two of the more promising wet technologies are absorption-reduction and oxidation-absorption-reduction.

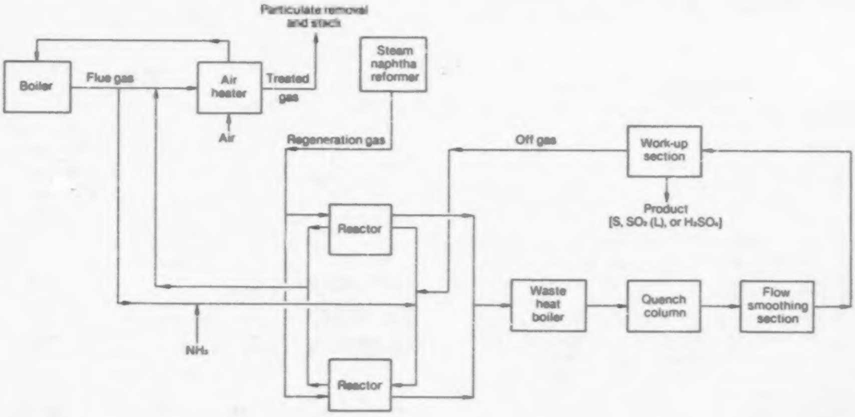


Fig. 8 — Flow diagram of the SFGT process.[15]

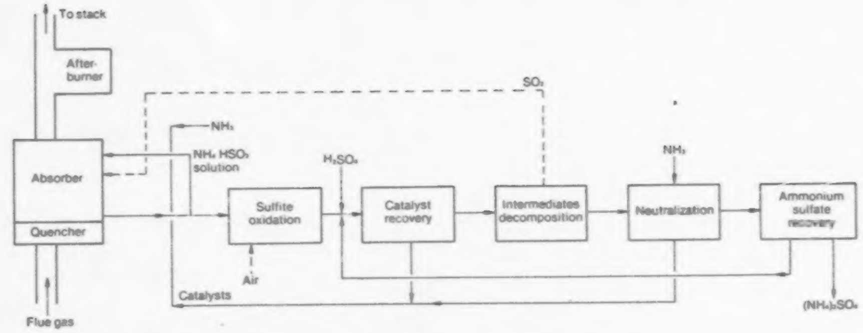


Fig. 9 — Block diagram of Chisso engineering process for the simultaneous removal of NO_x and SO_2.[9]

The absorption-reduction processes were developed to avoid gas phase oxidation by using ferrous chelating compounds Fe · EDTA to directly improve the absorption of NO by scrubbing solutions.[11] The NO is absorbed and forms a complex with the chelating compound per the following reaction.

$$NO + FE \cdot EDTA \rightarrow Fe \cdot EDTA \cdot NO$$

SO_x is simultaneously absorbed as sodium sulfite and reacts with the NO complex to yield nitrogen, a regenerated chelating compound and sodium sulfate as follows:

$$Na_2CO_3 + SO_2 \rightarrow Na_2SO_3$$

$$Na_2SO_3 + Fe \cdot EDTA \cdot NO \rightarrow$$
$$Fe \cdot EDTA + Na_2SO_4 + \tfrac{1}{2}N_2$$

These processes are capable of achieving 90% NO_x and SO_x reductions. These processes involve considerable equipment and have complex chemistry which make them less desirable then dry processes.

Chisso Engineering has a pilot scale process based on this technology. Chisso's process uses an ammoniacal solution containing EDTA and ferrous ion, and produces ammonium sulfate. A flow diagram of this process is presented in Fig. 9.

The oxidation-absorption-reduction processes are essentially modified FGD processes.[11] In these processes a gas phase oxidizing agent, normally ozone or chlorine dioxide (ClO_2), is injected into the flue gas stream ahead of the scrubber to convert NO to NO_2, by the following reactions:

$$NO + O_3 \rightarrow NO_2 + O_2$$

$$\text{or } 2NO + ClO_2 + H_2O \rightarrow NO_2 + HNO_3 + HCl$$

NO_2 is then absorbed into an aqueous solution containing sulfite and nitrates while SO_x is absorbed and forms the sulfite or bisulfite ion in the scrubbing solution as follows:

$$2NO_2 + 2CaSO_3 + 2CaCO_3 + 2H_2O \rightarrow$$
$$2Ca(NO_2)_2 + 2CaSO_4 \cdot 2H_2O + 2CO_2$$

$$CaSO_3 + SO_2 + H_2O \rightarrow Ca(HSO_3)_2$$

Sulfite ions are then partially oxidized to sulfate during reduction of the NO_2.

$$Ca(HSO_3)_2 + Ca(NO_2)_2 + H_2O \rightarrow$$
$$N_2 + 2CaSO_4 \cdot 2H_2O$$

As with many FGD processes[16] the ultimate disposal of the salts produced pose a real problem. Other problems include the complex chemistry and extensive equipment requirements.

These oxidation-absorption-reduction processes have the potential for 90% NO_x reduction and 95% SO_x reduction. A flow diagram of the Mitsubishi Heavy Ind. process is shown in Fig. 10.

COMMERCIAL APPLICATIONS

At present there are approximately 110 NO_x or NO_x/SO_x treating processes either operating or in some phase of startup. Of these 26 units (24%) are in the United States and the remaining units are in Japan. A listing of all applications by process type is presented in Table 6.

The most utilized technology is selective catalytic reduction with NH_3 which accounts for 70 units or 64% of the total. The main reasons for this leadership role are

TABLE 6 — Flue gas treating applications in the U.S. and Japan[9-10]

Process Type	For	Units	% of Total
Selective Catalytic Reduction	NO_x	70	63.7
Selective Reduction Exxon Thermal DeNOx	NO_x	25	22.7
Shell/UOP Selective Reduction-Adsorption	NO_x and SO_2	2	1.8
Wet Oxidation-Absorption Reduction	NO_x and SO_2	9	8.2
Wet Absorption-Reduction	NO_x and SO_2	4	3.6
	Total	110	100.0

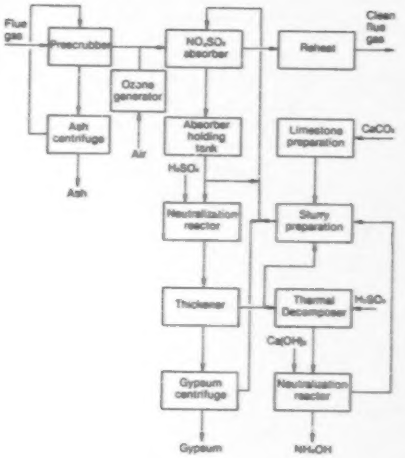


Fig. 10 — Process flow diagram for MHI oxidation-absorption-reduction process.[19]

low capital costs, low operating expense and reliability. Of this total 66 units or 94% are in Japan.

The most widely used technology in the U.S. is the Exxon Thermal DeNO$_x$ system with 21 applications. Most of these applications are retrofits on oil refinery process heaters in California. The main reason for this is low capital cost and short installation time while achieving the required NO$_x$ reduction. Only four units have been used in Japan. A listing of all U.S. FGT applications is presented in Table 7.

Due to higher capital costs and process complexity there are only 13 applications (11.8% of the total) of wet process technology, all of which are in Japan. These are summarized in Table 8. A listing of all other FGT applications in Japan is presented in Table 9.

TABLE 7 — Flue gas treating applications in the U.S.[8,12,17]

Process Type	No. of Units	Location	Supplier
SCR-NO$_x$	1	Georgia Power Company	Hitachi Zosen
SCR-NO$_x$ Ads-SO$_2$	1	Tampa Electric	Shell/UOP
SCR-NO$_x$	1	Public Service of Colorado	Kawasaki
SCR-NO$_x$	1	Southern California Edison	Kawasaki
SR-NO$_x$	1	L.A. Dept. of Water & Power	Exxon
SR-NO$_x$	20	California Refineries	Exxon
SCR-NO$_x$	1	U.S.A. Petroleum	Fallon Engineering

TABLE 8 — Wet NO$_x$/SO$_x$ control plants in Japan[19]

Process Developer	Capacity (Nm³/hr)	Source of Gas	Start-up
Oxidation/Absorption/Reduction processes			
Chiyoda	1,000	Oil-fired Boiler	Aug. 1973
Ishikawahima H.I.	5,000	Oil-fired Boiler	Sep. 1975
Mitsubishi H.I.	2,000	Oil-fired Boiler	Dec. 1974
Osaka Soda	60,000	Oil-fired Boiler	Mar. 1976
Shirogane	48,000	Oil-fired Boiler	Aug. 1974
Sumitomo Metal-Fujikasui	62,000	Oil-fired Boiler	Dec. 1973
Sumitomo Metal-Fujikasui	100,000	Heating Furnace	Dec. 1974
Sumitomo Metal-Fujikasui	39,000	Oil-fired Boiler	Dec. 1974
Sumitomo Metal-Fujikasui	25,000	Sintering Machine	— 1976
Absorption/Reduction processes			
Asahi Chemical	600	Oil-fired Boiler	Apr. 1974
Chisso Corp.	300	Oil-fired Boiler	Apr. 1974
Kureha Chemical	5,000	Oil-fired Boiler	Apr. 1975
Mitsui S.B.	150	Oil-fired Boiler	Apr. 1974

TABLE 9 — Selective catalytic reduction plants in Japan[10,12]

Process Developer	Gas Source	Fuel[a]	Capacity (Nm³/hr)	Type of Reactor[b]	Start-up
Asahi Glass	Furnace	—	70,000	IMB	—
Babcock-Hitachi	Boiler	Coal	—	—	Oct. 1980
Babcock Hitachi	Boiler	Coal	—	—	June 1981
Hitachi, Ltd.	Coke Oven	COG	500,000	IMB	Nov. 1976
Hitachi, Ltd.	Boiler	HO(HS)	15,000	IMB	Oct. 1977
Hitachi, Ltd.	Boiler	Kerosene	18,000	FB	Oct. 1977
Hitachi, Ltd.	Boiler	HO(LS)	20,000	FB	Aug. 1977
Hitachi, Ltd.	Boiler	HO(LS)	19,000	FB	July 1977
Hitachi, Ltd.	Boiler	Crude Oil	300000	FB	July 1977
Hitachi, Ltd.	Boiler	LNG	2,000,000x2	FB	Apr. 1978
Hitachi, Ltd.	Boiler	HO	20,000	IMB	Apr. 1978
Hitachi, Ltd.	Boiler	Kerosene	30,000	FB	Dec. 1978
Hitachi, Ltd.	Boiler	HO(LS)	480,000	PPC	June 1978
Hitachi, Ltd.	Boiler	HO(LS)	550,000	PPC	June 1978
Hitachi Zosen	Boiler	CO	350,000	FB	Oct. 1975
Hitachi Zosen	Boiler	HO(HS)	440,000	FB	Nov. 1975
Hitachi Zosen	Furnace	Kerosene	71,000	FB	May 1976
Hitachi Zosen	Furnace	Coke, Oil	782,000	FB	Nov. 1976
Hitachi Zosen	Furnace	HO	10,000	FB	Dec. 1976
Ishikawajima H.I.	Boiler	Crude Oil	20,000	HC	Apr. 1977
Ishikawajima H.I.	Boiler	HO(LS)	180,000	HC	Jan. 1978
Ishikawajima H.I.	Boiler	HO(LS)	980,000	HC	Apr. 1978
Ishikawajima H.I.	Boiler	HO(LS)	480,000	HC	June 1978
Ishikawajima H.I.	Boiler	Crude Oil	1,000,000	HC	Apr. 1979
Ishikawajima H.I.	Boiler	HO(LS)	1,800,000	HC	July 1979
Ishikawajima H.I.	Boiler	HO(LS)	1,880,000	HC	Aug. 1981
JGC	Furnace	HO(MS)	90,000	PPR	Nov. 1975
JGC	Boiler	CO	70,000	PPR	July 1976
JGC	Coke Oven	COG	180,000	PPR	Mar. 1977
Kobe Steel	Coke Oven	COG	104,000	IMB	Aug. 1977
Kurabo	Boiler	HO(HS)	30,000	CMB	Aug. 1975
Kawasaki H.I.	Boiler	Coal	—	—	June 1981
Mitsubishi H.I.	Boiler	HO(LS)	300,000	IMB	Sept. 1978
Mitsubishi H.I.	Boiler	LNG	15,000x2	FB	Dec. 1978
Mitsubishi H.I.	Boiler	Coal	—	—	Apr. 1980
Mitsubishi H.I.	Boiler	HO(LS)	40,000	PPC	Mar. 1977
Mitsubishi H.I.	Boiler	HO(LS)	200,000	HC	Jan. 1978
Mitsubishi H.I.	Boiler	LNG	1,880,000x2	FB	Oct. 1978
Mitsubishi H.I.	Boiler	HO(LS)	1,010,000	PPC	Feb. 1978
Mitsubishi H.I.	Boiler	HO(LS)	480,000	PPC	July 1978
Mitsubishi H.I.	Boiler	HO(LS)	1,820,000	HC	Feb. 1980
Mitsubishi Kakoki	Boiler	HO(HS)	14,000	FB	July 1976
Mitsubishi Kakoki	Furnace	Naphtha	19,000	FB	Oct. 1976
Mitsubishi Kakoki	Boiler	Kerosene	10,000	FB	Nov. 1976
Mitsubishi Kakoki	Boiler	Kerosene	30,000	FB	Oct. 1977
Mitsubishi Kakoki	Boiler	Kerosene	43,000	FB	Oct. 1979
Mitsui Engineering	Boiler	CO	200,000	FB	Oct. 1975
Mitsui Engineering	Boiler	HO(LS)	220,000	TC	Apr. 1976
Mitsui Toatsu	Furnace	Off Gas	87,000	FB	Feb. 1976
Mitsui Toatsu	Furnace	Off Gas	91,000	FB	Sept. 1976
Mitsui Toatsu	Furnace	Off Gas	170,000	FB	Jan. 1977
Mitsui Toatsu	Furnace	Off Gas	383,000	FB	June 1977
Mitsui Toatsu	Boiler	Off Gas	300,000	FB	Oct. 1977
Sumitomo Chemical Co.	Boiler	HO(HS)	30,000	FB	July 1973
Sumitomo Chemical Co.*	Furnace	LPG	200,000	FB	May 1974
Sumitomo Chemical Co.	Furnace	LPG	250,000	FB	Jan. 1975
Sumitomo Chemical Co.	Furnace	LPG	200,000	FB	Apr. 1975
Sumitomo Chemical Co.	Furnace	LPG	200,000	FB	Apr. 1975
Sumitomo Chemical Co.	Furnace	LPG	100,000	FB	—
Sumitomo Chemical Co.	Furnace	LPG	10,000	FB	—
Sumitomo Chemical Co.	Boiler	HO(MS)	240,000	FB	Mar. 1975
Sumitomo Chemical Co.	Boiler	HO(MS)	300,000	IMB	Oct. 1975
Sumitomo Chemical Eng.	Boiler	Naphtha	31,000x2	FB	Oct. 1977
Sumitomo Chemical Eng.	Boiler	Naphtha	23,000	FB	Dec. 1977
Sumitomo Chemical Eng.	Boiler	Naphtha	23,000	FB	June 1976
Sumitomo Chemical Eng.	Boiler	Naphtha	19,000	FB	July 1978

[a] Fuel Codes

CO	Carbon Monoxide
COG	Coke Oven Gas
HO	Heavy Oil
HO(HS)	High Sulfur Heavy Oil
HO(LS)	Low Sulfur Heavy Oil
HO(MS)	Medium Sulfur Heavy Oil
LNG	Liquefied Natural Gas
LPG	Liquefied Propane Gas

[b] Reactor Codes

CMB	Continuous Moving Bed
FB	Fixed Bed
HC	Honeycomb Catalyst
IMB	Intermittent Moving Bed
PPC	Parallel Plate Catalyst
PPR	Parallel Passage Reactor
TC	Tubular Catalyst

TABLE 10 — Cost of NO_x and NO_x/SO_x systems in the U.S.[9,11]

Process Type	Pollutant Removed	Capital Cost ($/kW)	Revenue Requirement (mills/kWh)
Selective Catalytic Reduction (SCR)	NO_x, Particulate	64	3.4
Flue Gas Desulfurization (FGD)	SO_2	101	4.2
SCR + FGD	NO_x, SO_2, Particulate	165	7.6
Dry, Simultaneous (UOP-Shell)	NO_x, SO_2, Particulate	160	7.3
Wet, Simultaneous	NO_x, SO_2, Particulate		
Moretana (ClO_2)		205	12.2
Asahi (EDTA)		233	12.6
IHI (Ozone)		482	19.8

Basis for the Estimate:

Particulate Control System	ESP for dry systems; wet scrubber for wet simultaneous systems
FGD System	Limestone
SO_2 Removal Efficiency	90%
NO_x Removal Efficiency	90%
Particulate Removal Efficiency	99.5%
Boiler Size	500 MW, new
Fuel	Coal
Heating value	24.4 MJ/kg (10,500 Btu/lb)
Sulfur content	3.5%
Ash content	16%
Operation	7000 hr/yr
Capital Investment	mid-1979
Annual Revenue Requirement	mid-1980

Economics of FGT Processes. For new boilers and process heaters combustion modifications are less expensive than flue gas treating and will continue to be the dominant control technology as long as NO_x emission limits can be met. For retrofit applications, especially on refinery or petrochemical process heaters, the Exxon Thermal $DeNO_x$ and SCR processes appear to be a cost-effective approach to meet new stringent NO_x regulations, such as in California.[7]

Actual investment and operating costs are dependent on many factors such as fuel type, unit size, percent removal required, new construction or retrofit, location, costs for chemicals (NH_3 or absorbing solutions) and catalyst.

One SCR application on a refinery process heater in California required a capital investment of $675,000.00.[17]

For coal fired boilers in the United States capital cost ranges from 64$/kW for SCR systems to 482$/kW for wet NO_x/SO_x systems, with revenue requirements for these processes ranging from 3.4 mills/kWh to 19.8 mills/kWh. These data are presented in Table 10.

Capital costs and revenue requirements for FGT processes in Japan are much lower, 15.5 to 111.1$/kW and 1.6 to 9.4 mills/kWh as listed in Table 11.

Capital cost requirements for the Exxon Thermal $DeNO_x$ process are approximately 1 cent per million Btu of capacity, with operating costs (mostly NH_3) of approximately 1.9 cents per million Btu.[13]

ACKNOWLEDGMENTS

The authors gratefully acknowledge the support and interest of Mr. W. D. "Dub" Haney, Jr. Opinions expressed in this paper are strictly those of the authors and do not necessarily reflect the official views of ARCO.

TABLE 11 — Cost of NO_x and NO_x/SO_x control systems in Japan[9]

Process Type	Pollutant Removed	Capital Cost ¥/kW	Capital Cost $/kW	Operating Costs ¥/kW	Operating Costs mills/kWh
Selective Catalytic Reduction (SCR)	NO_x	2800	15.5	0.3	1.6
Flue Gas Desulfurization (FGD)	SO_2	14000	77.8	1.2	6.7
SCR + FGD	NO_x & SO_2	18500	102.8	1.5	8.3
Simultaneous NO_x/SO_x (Dry and Wet)	NO_x & SO_2	20000	111.1	1.7	9.4

Basis for estimate:

Plant Size	300 MW, new	Maintenance	3% of investment cost
Fuel	oil	Insurance	2% of investment
NO_x Concentration	200 ppm	Overhead	5% of investment cost
SO_x Concentration	1500 ppm	Catalyst Life	2 years
Particulate Concentration	200 mg/Nm³	Annual Operation	8,000 hours
Temperature	360°C	Ammonia	¥80/kg
NO_x Removal Efficiency	80%	Power	¥12/kWh
SO_2 Removal Efficiency	90%	Steam	¥2/kg
Depreciation	7 years	Kerosene	¥32/kg
Interest Per Year	10%	Monetary Conversion Rate	¥180/$

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This selection, dealing with combustion modifications, is part of a Department of Commerce report on "NOx abatement for stationary sources in Japan." Modifications to alter the operating conditions and the combustion of fuels have proven to be simple, relatively inexpensive methods to reduce fuel and thermal NOx.

SECTION 2

NO_x ABATEMENT BY COMBUSTION MODIFICATION (CM)

2.1 CLASSIFICATION OF COMBUSTION MODIFICATION TECHNIQUES

NO_x emissions from combustion processes are formed by two mechanisms: the oxidation of nitrogen in air at high temperatures, and the combustion of various nitrogen compounds contained in fuel. The former is called "thermal NO_x," and the latter "fuel NO_x."

Formation of thermal NO_x can be reduced effectively by combustion modifications, such as decreasing the oxygen concentration in the combustion regions; shortening the residence time of combustion gases in high temperature zones; or lowering the flame temperature. Reduction of fuel NO_x emissions is achieved by decreasing the oxygen concentration in the combustion reaction zone and switching to a low-nitrogen fuel.

Combustion modification techniques used widely in Japan for NO_x control can be classified into four categories:

- Change of operating conditions
  1) Low-excess-air combustion
  2) Change in contacting of fuel with air in combustion chamber
  3) Reduction of heat load in combustion chamber (reduction of output power)
  4) Lowering of air preheating temperature

- Modification of burner design

- Modification of combustion system design
  1) Reduction of heat load in combustion chamber (increase of furnace volume)
  2) Staged combustion
  3) Flue gas recirculation
  4) Water or steam injection

- Other methods
  1) Change of fuel
  2) Modification of firing

Combinations of the above techniques are also in practical use. Although operating conditions can be changed relatively easily in existing installations, the changes usually reduce the NO_x emissions only slightly and often cause operating difficulties. On the other hand, modifications of the burner and combustion system design are promising control techniques. Fewer problems are encountered when design modifications are incorporated in new plants; relatively large reconstruction costs are required when the modifications are applied to existing installations.

The principles and characteristics of NO_x control techniques are described in the following section.

2.2 CHANGE OF OPERATING CONDITIONS

2.2.1 Low-Excess-Air Combustion (LEA)

Low-excess-air combustion, a means of preventing low-temperature corrosion caused by SO_3 in large boilers, is also useful in NO_x abatement. LEA combustion reduces the formation of thermal NO_x and fuel NO_x by decreasing the oxygen concentration in combustion regions. There it has been applied to large boilers burning high-nitrogen fuels, such as grade B (0.08-0.35 percent N) or grade C (0.20-0.35 percent N) heavy oil. On the other hand, the technique is not widely used with small boilers. The reasons are:

- Small boilers usually burn low-nitrogen fuels such as kerosene, grade A heavy oil (0.005-0.08 percent N), or gas. Thus, there is no need to attempt fuel NO_x reduction.
- Small boilers normally are unmanned, and their operation is not very well controlled. Therefore, operation with low excess air in these boilers tends to cause incomplete combustion.

Recently, LEA combustion has been applied to large metal heating furnaces as a means of NO_x abatement and fuel conservation.[1]

2.2.2 Improvement of Fuel/Air Mixing Conditions in Combustion Chamber

It has been reported that NO_x is reduced considerably through a slight adjustment of the air register, for example, a change in the vane angle. The reduction may be due to the change in mixing condition in the combustion chamber. Few existing installations use this technique.

Reduction of NO_x by a change in mixing conditions is achieved either with very good or poor mixing. With good mixing, the combustion rate increases and temperature in the reaction zone becomes high. Yet thermal NO_x formation can be reduced because the residence time in the high temperature region can be shortened. When the mixing condition is poor, the combustion rate slows down and the flame temperature drops. This results in the reduction of both thermal and fuel NO_x emissions in the same way as in the staged-combustion type burner. (The latter is discussed in section 2.3.5.)

Changing the air system makes it necessary to readjust the air/fuel ratio. An improper adjustment may increase the formation of unburned products such as soot, carbon monoxide, and hydrocarbons, or may cause unstable combustion.

2.2.3 Reduction of Heat Release Rate in the Combustion Chamber (Reduction of Output Power)

Generally, when the heat load in the combustion chamber decreases, the flame temperature is lowered and thermal NO_x emissions are reduced. The use of this method in existing installations, however, results in a decrease in fuel efficiency and output power. Therefore, this method is regarded only as an emergency measure.

2.2.4 Reduction of Air Preheating

Lowering of the air preheating temperature reduces the flame temperature and consequently the formation of thermal NO_x. This method is seldom used in existing installations because it also reduces fuel efficiency. Moreover, its application is limited to large boilers with air preheaters, for which better control techniques are available.

2.3 MODIFICATION OF BURNER DESIGN

2.3.1 Introduction

This method reduces NO_x with the use of an improved burner, the so-called low-NO_x burner (LNB). Low-NO_x burners developed in Japan are mostly applied to boilers and various kinds of furnaces. They may be classified into five types:

- Good-mixing
- Divided-flame

- Self-recirculation
- Staged-combustion
- Combination

Large low-NO_x burners developed so far have capacities of 5×10^7 kcal/hr for boilers and 2×10^7 kcal/hr for heating furnaces which are large enough for commercial use.

2.3.2 Atomizing Type Burners

Atomizing type low-NO_x burners are useful in reducing the formation of thermal NO_x by shortening the residence time of combustion gas in high temperature regions. They have little effect on reducing fuel NO_x, however, because of the high oxygen concentration in the initial stage of combustion.

An example of an atomizing type low-NO_x burner is illustrated in Figure 2-1. This burner, developed by Nippon Furnace Kogyo Kaisha Ltd., is called the NFK-TRW burner because it has been commercialized under a technical license agreement with Civiltech Corp. (U.S.) in a joint venture with TRW Inc.

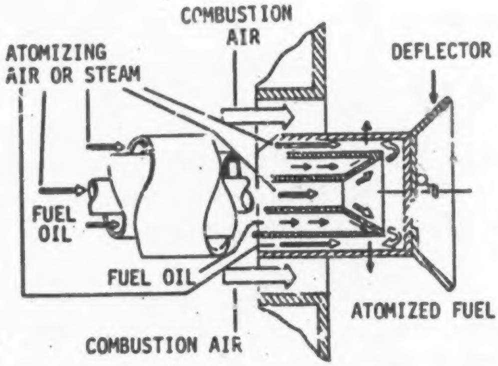
COMBUSTION AIR
ATOMIZING AIR OR STEAM
DEFLECTOR
FUEL OIL
FUEL OIL
ATOMIZED FUEL
COMBUSTION AIR

Figure 2-1. NFK-TRW Burner.

In this burner, atomizing of fuel with air is achieved by impinging the air stream at nearly right angles against the jets of fuel injected radially outward through a large number of shaped ports. The air and fuel are then further mixed by means of a deflector, which functions as a flame holder. The mixture of fuel and air produces a thin, conical flame according to the flow pattern. Due to the concical shape, a maximum radiation surface is produced, and heat from the flame is rapidly removed. Therefore, the residence time of combustion gas in the high temperature regions is shortened, and thermal NO_x is reduced. The NO_x concentrations produced using the burner at various operating conditions are shown in Figure 2-2. Burners which can burn either oil or gas have been commercialized. They reduce not only NO_x, but also soot emissions. In addition, they allow the heat load in the combustion chamber to be maintained as high as before, thereby increasing fuel efficiency. These burners are applicable only to boilers, however, because of their flame shape. The pressure drop at the burner throat is relatively high.

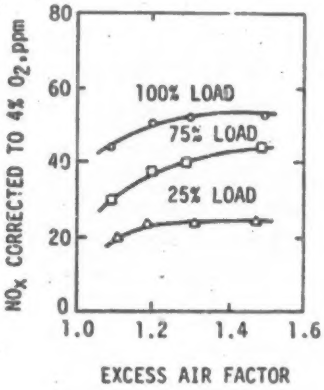

Figure 2-2. Effect of NFK-TRW Burner on NO_x. (Fuel, grade A fuel oil; air temperature, 25°C; maximum fuel flow rate, 380 ℓ/hr)

2.3.3 Divided-Flame Type Burners

In the divided-flame type low-NO_x burner for heavy oil, the flame is divided into several independent small flames by means of grooved nozzle tips. This prevents mutual interference between small flames. Hence, the flame temperature and the residence time in the high temperature regions decrease, resulting in thermal NO_x reduction.

The burner, developed by Ishikawajima-Harima Heavy Industries Ltd. (IHI), is shown in Figure 2-3. Its effect when combined with other NO_x control techniques is illustrated in Figure 2-4.[3] The two curves indicate the practical upper limit and lower limit. NO_x reduction with this burner decreases with an increase in the effects of other control techniques. For example, the reduction ratio due to the effect of the burner ranges from 18-42 percent without other CM techniques and from about 10-30 percent when 40 percent reduction is achieved by other techniques.

2.3.4 Self-Recirculation Type Burners

The configuration of the self-recirculation type burner is shown in Figure 2-5. This burner has been developed by Daido Tokushuko Co. under a technical license agreement with Caloric Gesellschaft für Apparate Bau (West Germany).[4]

It reduces both thermal and fuel NO_x formation in such a manner that partially combusted hot gas is reintroduced into the initial stage of combustion by the fuel jet and atomizing air or steam. Thus, both vaporization of fuel and firing in the low-oxygen-concentration zone are accomplished in the initial stage of combustion.

This burner's effect on NO_x reduction is illustrated in Figure 2-6. This type of burner is effective not only for NO_x reduction, but also for soot suppression by hot combustion gas recirculation. It is used in metal heating furnaces that burn oil and gas.

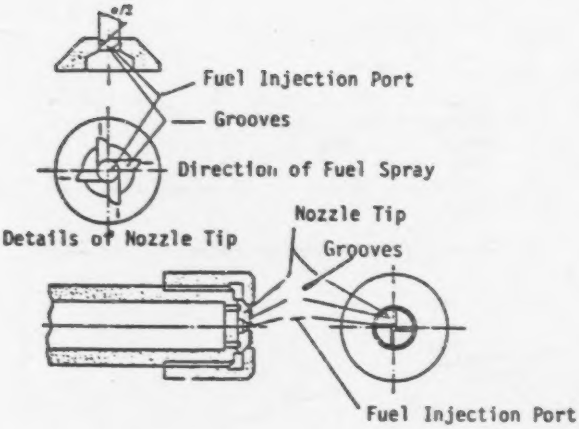
θ/2
Fuel Injection Port
Grooves
Direction of Fuel Spray
Details of Nozzle Tip
Nozzle Tip
Grooves
Fuel Injection Port

Figure 2-3. Divided-flame type burner.

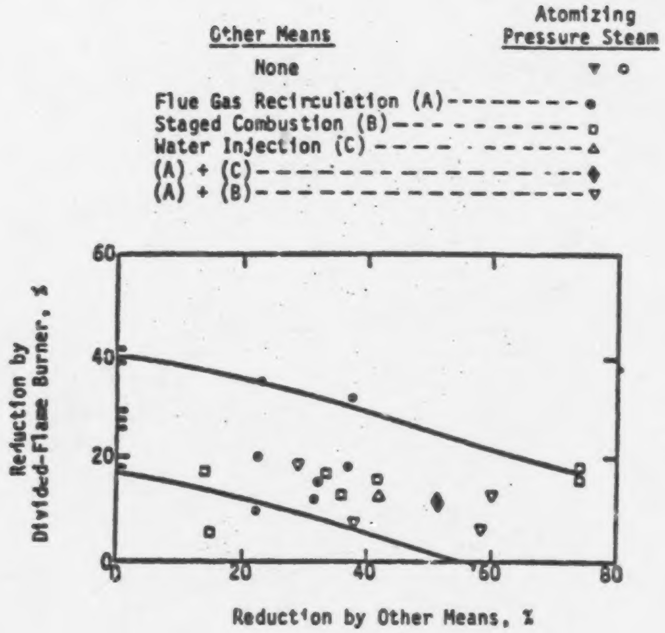

Figure 2-4. NO$_x$ reduction by divided-flame type burner and other means.

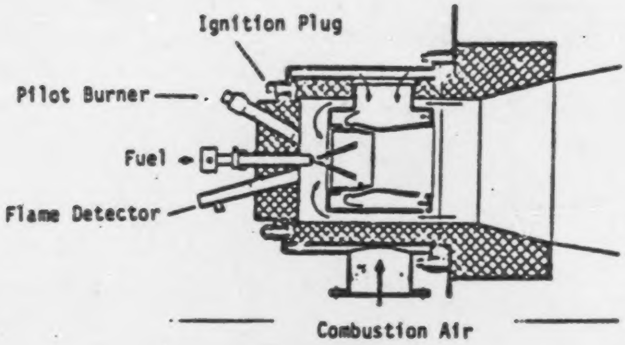
Ignition Plug
Pilot Burner
Fuel
Flame Detector
Combustion Air

Figure 2-5. Self-recirculation type burner.

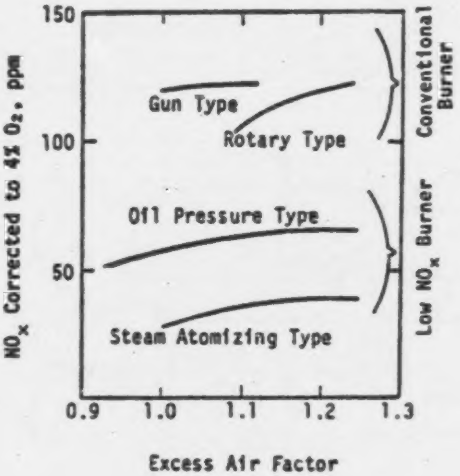

Figure 2-6. Effect of self-recirculation type burner on NO_x reduction. (Fuel, grade A oil; Fuel flow rate, 100 ℓ/hr)

A burner designed exclusively for gas firing is shown in Figure 2-7. This burner developed by Sanray Reinetsu Co., Ltd., is called the XB burner. Several fuel gas injection nozzles are installed circularly to shorten the flame. NO_x levels are presented in Figure 2-8. A similar type of burner has been developed by Chugai Ro Kogyo Co., Ltd.

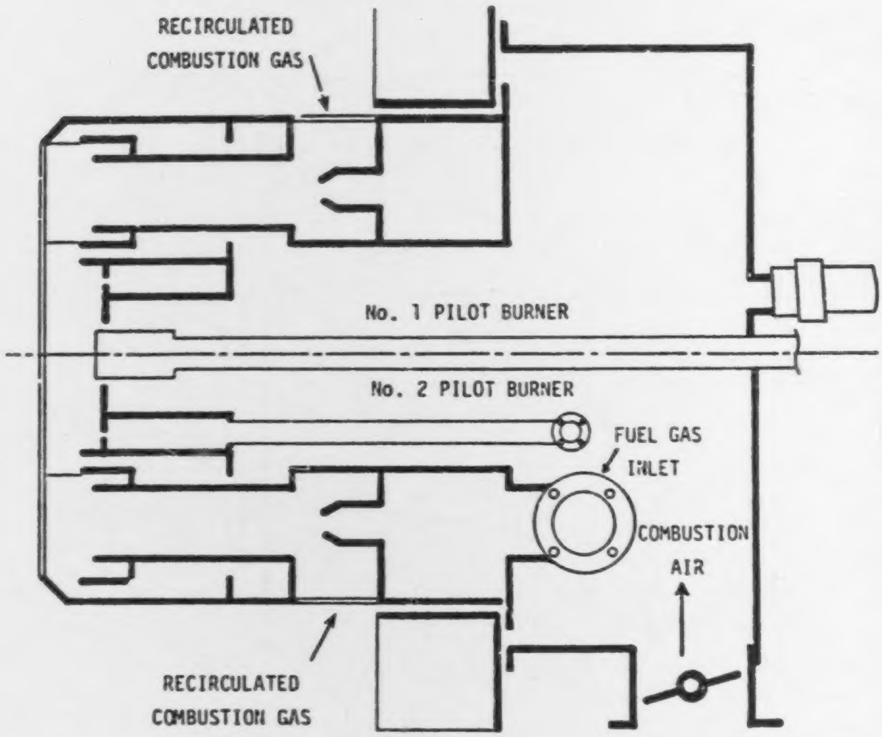
RECIRCULATED COMBUSTION GAS
No. 1 PILOT BURNER
No. 2 PILOT BURNER
FUEL GAS INLET
COMBUSTION AIR
RECIRCULATED COMBUSTION GAS

Figure 2-7. XB Burner.

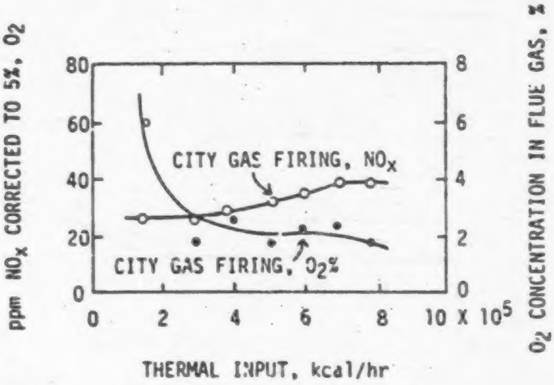

Figure 2-8. Effect of XB burner on NO_x emission levels.

2.3.5 Staged-Combustion Type Burners

Staged-combustion type burners are divided roughly into a two-stage combustion group and an off-stoichiometric combustion group. Burners in the two-stage combustion group are classified according to the equivalence ratio in the first stage into fuel-rich and air-rich. The fuel-rich type is used usually for two-stage combustion, and the air is added in the second stage; fuel-rich type burners are shown in Figures 2-9, 2-11, and 2-13. In the air-rich type, fuel is injected in the second stage; an example of this type of burner is presented in Figure 2-15.

Figure 2-9 illustrates the APOC burner developed by Tokyo Gas Company for firing town gas.[5] A more advanced burner of this type has a catalyst at the outlet of the preliminary combustion region for further reduction of NO_x. Figure 2-10 shows the emission levels with this burner. HCN and NH_3 are likely to play a significant role in NO_x formation in the second stage. The catalyst decreases this problem by reducing HCN and NH_3 together with NO_x in the first stage. The advanced type is not yet used commercially.

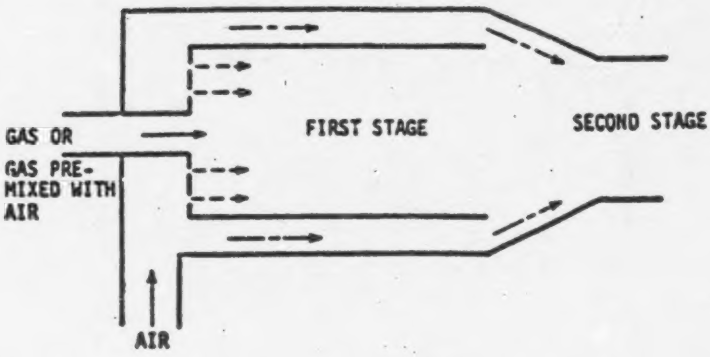
GAS OR
GAS PRE-
MIXED WITH
AIR
AIR
FIRST STAGE
SECOND STAGE

Figure 2-9. APOC burner.

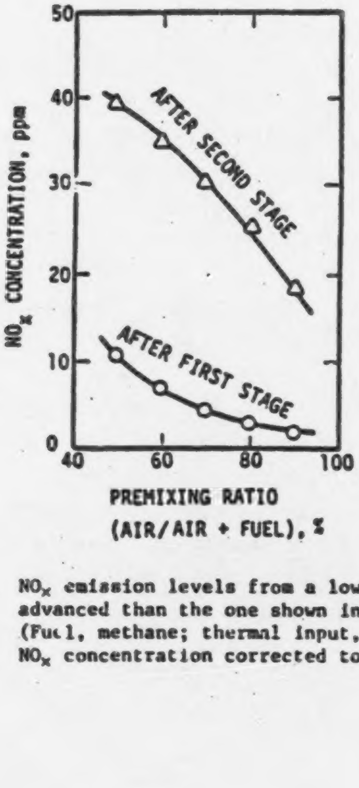

Figure 2-10. NO_x emission levels from a low-NO_x burner more advanced than the one shown in Figure 2-9. (Fuel, methane; thermal input, 95,000 kcal/hr; NO_x concentration corrected to 0 percent O_2)

The FH burner illustrated in Figure 2-11, used for LPG, heavy oil or kerosene fuels, is also of the fuel-rich type.[6] Figure 2-12 shows its effect on NO_x reduction. This burner, developed by Chugai Ro Kogyo Co., Ltd., is used in the heating furnace. Similar types of burners have been developed by many manufacturers because of their relatively simple configuration and the capability of fuel NO_x reduction. These manufacturers include Mitsubishi Heavy Industries, Ltd., Kawasaki Heavy Industries, Ltd., Osaka Gas Company, Sanray Reinetsu Co., Ltd. and Rozai Kogyo Co., Ltd.

The fuel-rich type burner shown in Figure 2-13 was designed for use with pulverized coal by Babcock Hitachi Co., Ltd.; it is called the dual-air register type burner and is used in utility boilers. This burner is effective for NO_x reduction, as shown in Figure 2-14. IHI has developed a similar type of burner.

An air-rich type two-stage combustion burner is shown in Figure 2-15.[8] This burner, developed by Tokyo Gas Company for gas firing, is called the TGS burner. The effect of the TGS burner on NO_x reduction is illustrated in Figure 2-16. In the first stage, the flame temperature is decreased due to high-excess-air combustion; in the second stage, the oxygen concentration is lowered due to low-excess-air combustion. These decreases in flame temperature and oxygen concentration provide NO_x reduction. It is unlikely, however, that complete combustion can be accomplished for oil or coal firing.

The basic structure of the off-stoichiometric combustion (OSC) type burner for oil firing is presented in Figure 2-17.[9] This burner has been developed by MHI for boilers using grade A heavy oil or kerosene. The fuel sprayed in the upper and lower compartments is evaporated by preheated air and premixed with air resulting in combustion. In the middle compartment, a normal diffusion flame is produced. Thus, off-stoichiometric combustion is realized. To reduce NO_x further, recirculated flue gas is mixed with air. MHI guarantees less than 70 ppm of NO_x with this system (see Figure 2-18).

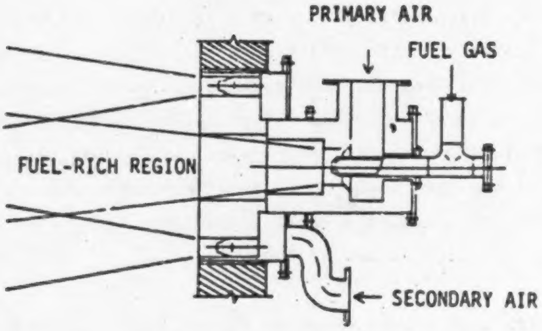
PRIMARY AIR
FUEL GAS
FUEL-RICH REGION
SECONDARY AIR

Figure 2-11. Schematic of FH burner.

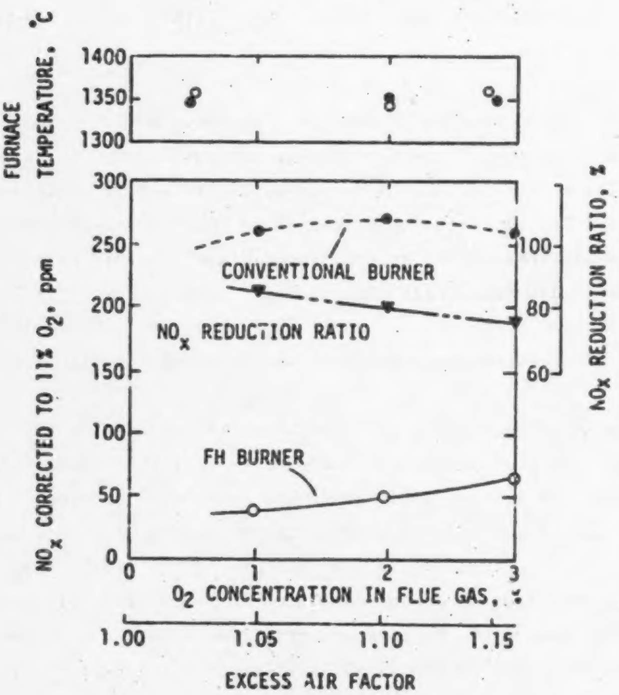

Figure 2-12. FH burner's effect on NO$_x$ reduction.

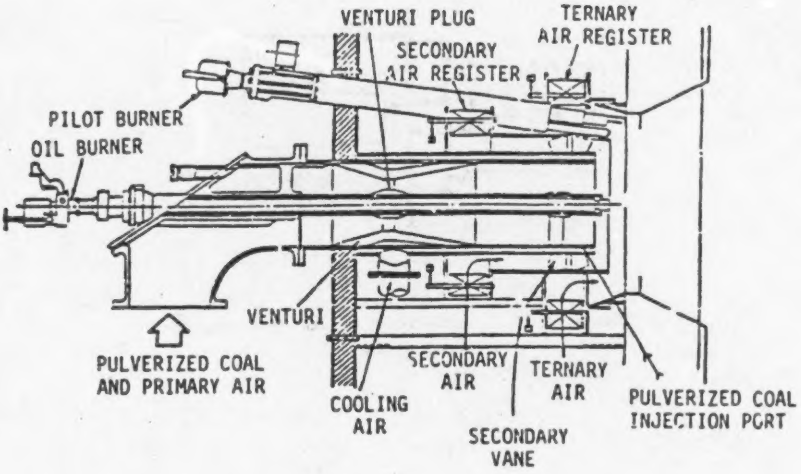
VENTURI PLUG
TERNARY AIR REGISTER
SECONDARY AIR REGISTER
PILOT BURNER
OIL BURNER
VENTURI
PULVERIZED COAL AND PRIMARY AIR
SECONDARY AIR
TERNARY AIR
COOLING AIR
SECONDARY VANE
PULVERIZED COAL INJECTION PORT

Figure 2-13. Dual-air register type burner for use with pulverized coal.

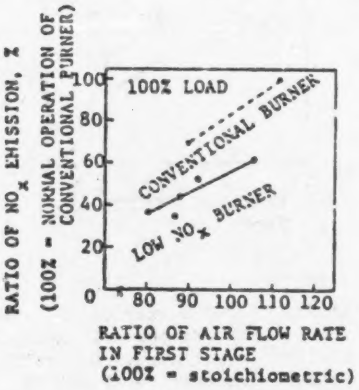

Figure 2-14. Effect of low-NO$_x$ burner shown in Figure 2-13. (Boiler capacity, 310 tons/hr; nitrogen content in fuel, 0.89-1.21 percent)

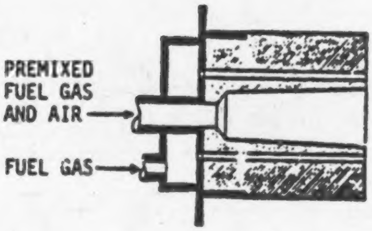
PREMIXED
FUEL GAS
AND AIR
FUEL GAS

Figure 2-15. Schematic of TGS burner.

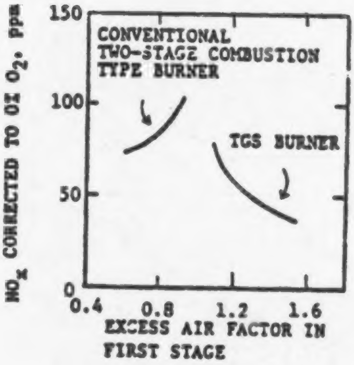

Figure 2-16. Effect of TGS burner on NO_x reduction. (Fuel, methane; thermal input, 2×10^5 kcal/hr; total excess air factor, 1.10; furnace temperature, 1,200°C)

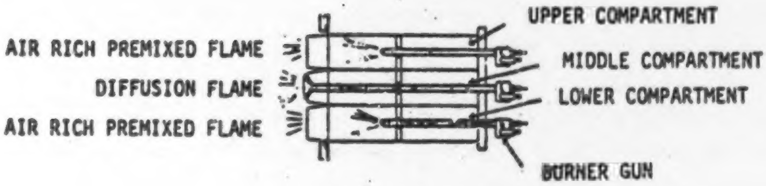
UPPER COMPARTMENT
AIR RICH PREMIXED FLAME
MIDDLE COMPARTMENT
DIFFUSION FLAME
LOWER COMPARTMENT
AIR RICH PREMIXED FLAME
BURNER GUN

Figure 2-17. Basic structure of OSC type burner.

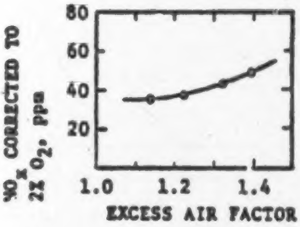

Figure 2-18. NO_x emission levels of the burner shown in Figure 2-17 with propane. (Air temperature, 250°C; thermal input, 3.5×10^7 kcal/hr; flue gas recirculation ratio, 15 percent)

Figure 2-19 shows the arrangement of the fuel injection holes of atomizing nozzles in an off-stoichiometric combustion burner for oil firing; this OSC burner was developed by Volcano Company.[10] The main feature of its atomizer is that not all of the fuel-injection holes are of the same size. As combustion air is uniformly admitted around the atomizer, the fuel-injection holes with larger diameters produce fuel-rich combustion zones, whereas the smaller holes produce fuel-lean regions. Therefore, off-stoichiometric combustion is achieved. Since several different arrangements of fuel-injection holes are available, the optimum arrangement may be determined by experimentation. Because of the ease of installation and the low cost, this burner has been used widely. NO_x levels produced by such a burner are presented in Figure 2-20. This burner tends to increase soot emissions when excess air is low, because some of the flame zones are deficient in oxygen. A similar type of burner has been developed by Babcock Hitachi Co., Ltd.

2.3.6 Combination Type Burners

There are several low-NO_x burners which combine the above-mentioned methods. For example, the burner illustrated in Figure 2-21 combines two-stage combustion and self-recirculation of hot combustion gas. It is called the SRG burner, and was developed by Nippon Furnace Kogyo Kaisha, Ltd.[2] This burner is adopted mainly in oil heating furnaces and boilers. The effect of the burner on NO_x reduction is shown in Figure 2-22. Figure 2-23 illustrates a similar type of burner (the TCG) developed by Osaka Gas Company.[11]

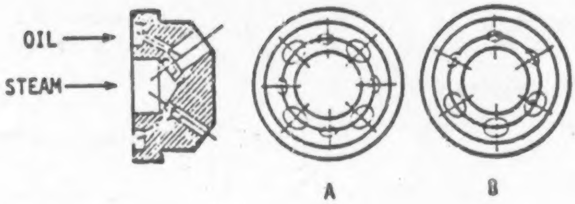

Figure 2-19. Atomizing nozzles in off-stoichiometric combustion-type low-NO_x burner for oil.

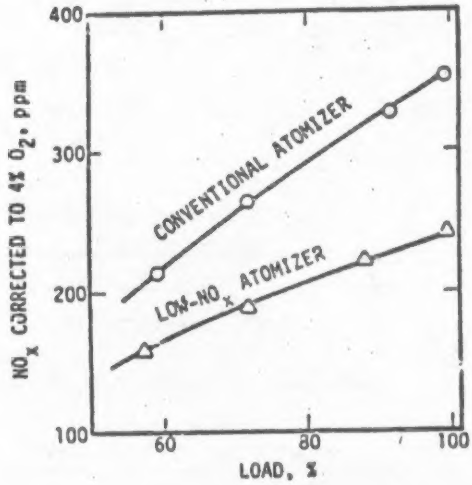

Figure 2-20. Effect of low-NO_x atomizer shown in Figure 2-19 on NO_x emissions. (Boiler capacity, 55 t/hr; fuel, grade C heavy oil; air temperature, 280°C)

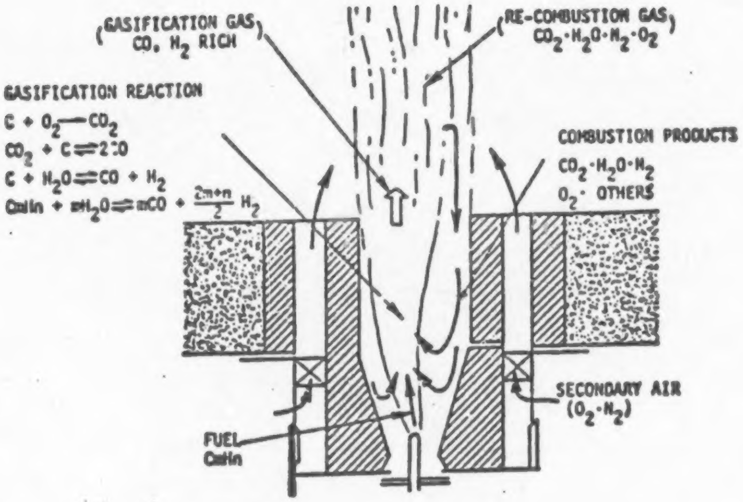
GASIFICATION GAS
CO, H_2 RICH
RE-COMBUSTION GAS
$CO_2 \cdot H_2O \cdot N_2 \cdot O_2$
GASIFICATION REACTION
$C + O_2 \longrightarrow CO_2$
$CO_2 + C \rightleftharpoons 2CO$
$C + H_2O \rightleftharpoons CO + H_2$
$C_mH_n + mH_2O \rightleftharpoons mCO + \frac{2m+n}{2} H_2$
COMBUSTION PRODUCTS
$CO_2 \cdot H_2O \cdot N_2$
$O_2 \cdot$ OTHERS
SECONDARY AIR
$(O_2 \cdot N_2)$
FUEL
C_mH_n

Figure 2-21. Schematic of SRG burner.

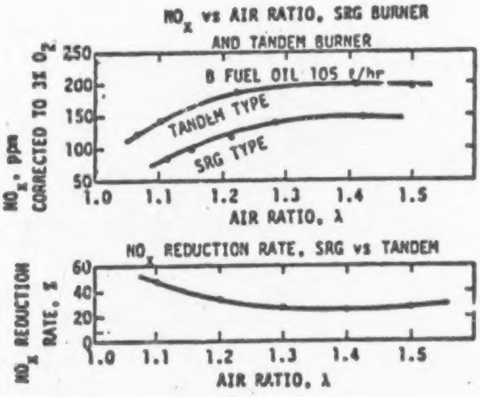

Figure 2-22. Effect of SRG burner on NO_x reduction.

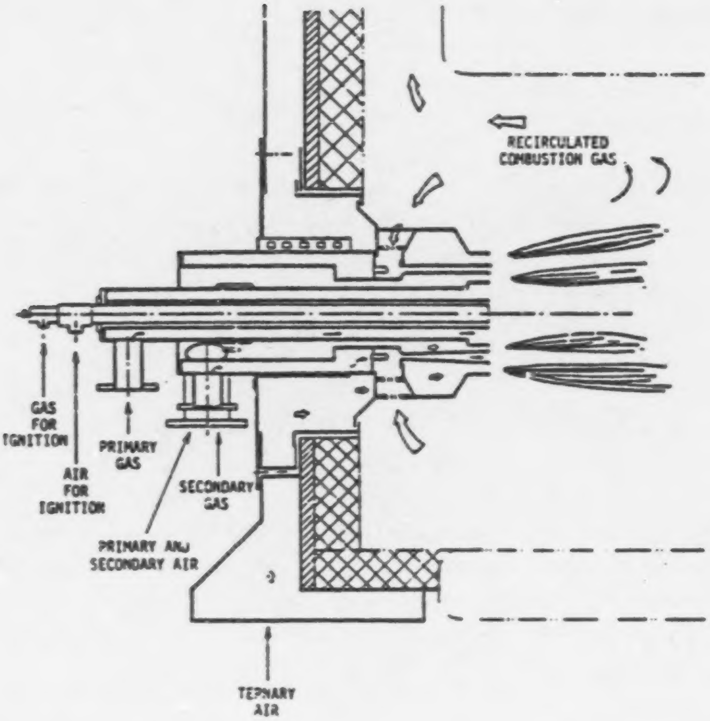
RECIRCULATED COMBUSTION GAS
GAS FOR IGNITION
AIR FOR IGNITION
PRIMARY GAS
SECONDARY GAS
PRIMARY AND SECONDARY AIR
TERNARY AIR

Figure 2-23. TCG burner.

2.4 MODIFICATION OF COMBUSTION SYSTEM DESIGN

2.4.1 Reduction of Heat Release Rate in the Combustion Chamber (Increase of Furnace Volume)

Manufacturers currently are designing boilers and furnaces with heat release rates about 10 to 20 percent lower than those of conventional ones. The new boilers and furnaces are larger and more easily adaptable to NO_x control techniques.

2.4.2 Staged Combustion

There are two major categories of staged combustion:

1) Two stage combustion (TSC)*

In two-stage combustion, about 70 to 90 percent of the stoichiometric air is admitted into the first stage. The additional air is then added downstream to obtain complete combustion. With this technique, both thermal and fuel NO_x are reduced by fuel-rich (low-oxygen concentration) combustion in the first stage.

Two stage combustion is classified into four types, depending on the locations of second-stage air ports:

a) On the furnace wall above the burners.
b) Top burners for air injection only.
c) Side or rear walls of the furnace.
d) On the circumference of the burners.

Large boilers installed in power plants employ type a or b. Type b, called quasi-two-stage combustion, is used in units to which type a cannot be applied. Types c and d are used in medium- and small-capacity boilers.

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*The number of stages seldom exceeds two.

Type c is not popular for retrofitting water tube boilers because it requires considerable revisions. Type d, which is illustrated in Figure 2-24, can be readily applied to small installations with a single burner because it calls for minor revisions. Two-stage combustion cannot be applied to installations with furnace dimensions that cannot accommodate greater flame lengths. Also two-stage combustion tends to increase the amount of unburned products; thus the unconsumed air increases in the flue as with normal combustion. When the air injection ports in the second stage are suitably located, further NO_x reduction is obtained.[12]

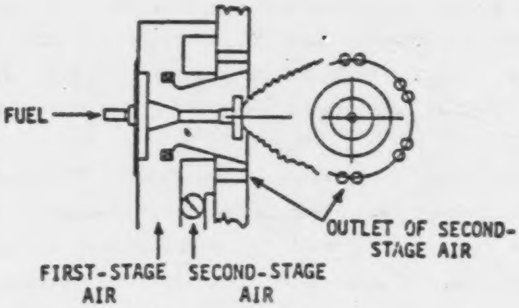

Figure 2-24. Two-stage combustion for small boiler.

2) Off-stoichiometric combustion (OSC)

This method has an effect similar to that of two-stage combustion; it is readily adapted to medium- and small-size boilers with several burners, to which two-stage combustion is not easily applied.

In this technique, the fuel-rich burners and air rich burners or air ports are suitably located. The fuel-rich portions reduce the oxygen concentration, and the air-rich portions decrease the temperature by means of the excess air; thus, NO_x formation is reduced. Combustion in the fuel-rich portions is completed upon mixing due to the excess air in the air-rich portions.

The location of fuel-rich and air-rich burners or air ports is decided after a systematic test. Often, it is effective to place the air-rich burners or air ports in the central upper parts of the furnace walls, or in the regions of highest heat release.

2.4.3 Flue Gas Recirculation (FGR)

In flue gas recirculation, part of the flue gas is mixed with combustion air and sent into the furnace. Thus, NO_x is reduced through the decrease of flame temperature resulting from the increase of the total heat capacity of combustion gas. Thermal NO_x formation can therefore be reduced.

Recirculation ratios are limited to about 30 to 40 percent to prevent unstable firing, although this limit is lower with larger units. Keeping the recirculation ratios in suitable ranges may improve combustion conditions and decrease the soot formation. However, the decrease of flame temperature alters the distribution of heat transfer and lowers the fuel efficiency of existing installations.

A recirculation fan and additional duct work are required to implement flue gas recirculation. As a result, installation cost is considerably higher, and more installation space is required than for the other methods. Therefore, this method is not used with small boilers or furnaces.

2.4.4 Water or Steam Injection

Water or steam injection reduces thermal NO_x emissions mainly by decreasing the flame temperature. There are two injection methods:

1) Injection into the combustion chamber. (This includes increasing the steam flow rate in the atomizer.)

2) Mixing water with the fuel (emulsification).

Injection ports close to the burners are effective for the first method. Thus, burners with water injection ports are adopted. The upper limit of the injection rate is about 5 kg/10^6 kcal.

With the use of this technique, formation of soot decreases, so it is possible to reduce the excess air. Therefore, the decrease in thermal efficiency may not be great.

There are two kinds of emulsified fuel: water droplets dispersed in oil (W/O type) and oil dispersed in water (O/W type). The former type can be used for NO_x reducing fuel because the resulting emulsion viscosity is lower than that of the O/W type reducing the pumping power requirements.

With the use of emulsified fuel oil, the atomized oil droplets are separated into very fine particles by sudden expansion of water contained in the oil droplets during the initial stage of combustion; this increases the contact area of fuel with air and also increases the combustion speed. These facts are likely to contribute to the reduction of thermal NO_x formation. Since water or steam is effective for suppression of soot formation, excess air can be reduced, and the decrease in thermal efficiency is not great. The increase of water concentration in the flue gas is small, so that its influence on metal corrosion can be neglected.

In the fuel emulsion method, usually a slight amount of emulsifying agent is used. Recently, a method that requires no agent has been developed through improvement of the mixer.

The price of a fuel emulsion system is about 50 x 10^5 to 60 x 10^5 yen (\$0.25 x 10^5 to \$0.30 x 10^5) for a capacity of 1 kl/hr, and the cost of an emulsifying agent is about 100 yen (\$0.50) per kiloliter.

Since it does not reduce fuel NO_x, this method is useful with low nitrogen oils.

2.5 OTHER METHODS

2.5.1 Change of Fuel

Except for special cases, NO_x concentrations are related to the nitrogen content of the fuel, which decreases in the following order: solid fuels (coal and coke), liquid fuels (petroleum fuel oils), and gaseous fuels (town gas, LNG and LPG).

Emissions from gas burning decrease in the following order: LPG (C_3H_8 and C_4H_{10}), LNG (CH_4), and town gas (synthetic gas with a heating value of about 5,000 kcal/Nm^3).

Hydrodesulfurization of heavy oil, which reduces sulfur content from 2-4 percent to 0.1-0.7 percent, also removes 20-30 percent of nitrogen, which is usually present in a concentration of 0.2-0.3 percent in the oil.

2.5.2 Modification of Firing

It is well-known that tangential firing systems give lower NO_x emissions than front and opposed firing systems, because the flame temperature is lower. This lower temperature is caused by better heat transfer resulting from a larger flame volume.

This Environmental Protection Agency document describes a variety of advanced combustion modification systems. Included in the discussion are a variety of distributed mixing systems, such as wall-fired and tangentially-fired boilers, as well as limestone injection/multi-staged burners (LIMB) and others.

ADVANCED COMBUSTION MODIFICATION

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ABSTRACT

Stationary combustion sources, particularly those fired with coal and oil, are major sources of NO_x and SO_x emissions. The commercially available combustion modification technology is relatively low cost; however, it gives a limited degree of control. Advanced combustion modification techniques are being developed which have the potential for greater NO_x control effectiveness. The development of new or improved combustion systems is based on an understanding of the basic processes of NO_x formation and control. This paper describes recent developments of advanced combustion techniques for control of NO_x and/or SO_x for coal, oil, and gas fired systems.

INTRODUCTION

Since the potential need for NO_x control technology in the U.S. was identified in the mid-1960's, a variety of options have been examined theoretically, experimentally, and in practical systems. For stationary combustion systems it has been generally accepted that the most cost effective technique is combustion modification. For commercial systems emphasis has been placed on external combustion modification techniques, such as flue gas recirculation and staged combustion, and on burner design. In general it has been found that flue gas recirculation is effective for gaseous fuels and light liquid fuels, whereas, staged combustion and burner design are effective for all fuels. While conventional combustion modifications are relatively inexpensive, their effectiveness is typically limited to 30 to 50 percent NO_x reduction. In an attempt to improve the effectiveness of combustion modification techniques, a number of advanced concepts are under development which have the potential for 80 to 90 percent reduction of emissions from uncontrolled levels. The purpose of this paper is to discuss the status of a number of advanced concepts developed under the U.S. EPA's NO_x Control Technology Program. In addition recent attempts to reduce SO_x emissions by sorbent addition with low NO_x firing systems will be described.

BACKGROUND

The basis for development of advanced combustion modification techniques is a basic understanding of the mechanisms of NO_x formation and control. The two sources of NO_x are: 1) formation of thermal NO_x by high temperature fixation of molecular nitrogen from the combustion air; and 2) conversion of nitrogen chemically bound in the fuel to form fuel NO_x. Combustion modification techniques are divided into two broad classes: 1) temperature reduction to control the formation of thermal NO_x; and 2) reduction of oxygen availability in the combustion zone to minimize conversion of fuel nitrogen to NO_x. While both conventional and advanced combustion modification techniques are based on the same broad classes, the optimization of the advanced techniques depends on a further understanding of two additional phenomena: 1) the effects of fuel composition on nitrogen evolution and reaction; and 2) the effects of combustion process parameters, such as air/fuel mixing history and heat extraction, on pollutant formation. The advanced processes represent optimization of the combustion system to achieve minimum NO_x emissions as discussed below.

Fuel Effects

The effects of fuel composition on NO_x formation and control are dominant for high nitrogen liquid and solid fuels; for gaseous and low nitrogen liquids, fuel properties have only minor effects. As a result, the development of advanced combustion techniques draws heavily on data related to fuel effects. Since the effects for liquid fuels and coal are significantly different, each fuel is discussed separately.

Coal. Until recently no systematic evaluation of the effects of coal properties on NO_x formation and control has been available; however, Chen (Ref. 1) has investigated 50 coals from 5 continents under a variety of combustion conditions. The NO_x emissions from a 21 thermal kilowatt (70,000 Btu/hr) experimental system using a burner which rapidly mixes the coal and combustion air at fuel-lean conditions are shown in Figure 1. (Note: Only NO is shown since NO_2 is a small fraction of the NO_x.) The major points to note are:

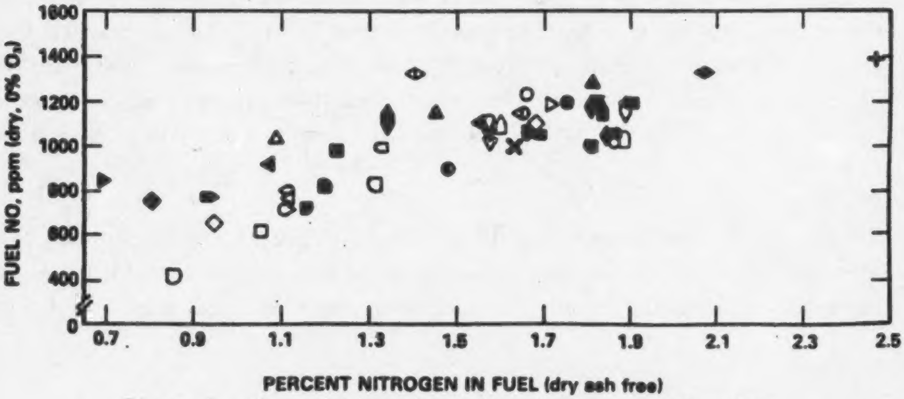

Figure 1. NO_x emissions from pulverized coal combustion under excess air conditions (Ref. 1)

1. NO_x emissions generally increase with increased fuel nitrogen; however,
2. For coals of similar rank and nitrogen content, the NO_x emissions can be significantly different.

He also found that the emissions correlated quite well with nitrogen evolution data from inert pyrolysis data.

The NO_x for staged combustion where the fuel-rich first stage was optimized for minimum NO_x emissions showed similar overall trends, with the absolute levels being reduced between 60 to 75 percent. He also showed that the minimum emissions were affected by the amount and type of volatile nitrogen species existing in the first stage and on the amount of nitrogen left in the char.

These results show the potential for significant effects of coal properties on the NO_x levels achievable for advanced systems. They also indicate the need for systems which drive the bulk of the nitrogen into reduced volatile nitrogen species in the first stage.

Liquid Fuels. Most petroleum derived liquid fuels have some fuel nitrogen content; however, the nitrogen tends to concentrate in the residual fuel oils, which may contain more than 0.8 percent by weight. Alternative liquid fuel derived from shale- and coal-derived crudes may have over 2 percent nitrogen by weight, and the nitrogen is more uniformly distributed through the product spectrum. The emission characteristics of a variety of petroleum-, shale-, and coal-derived liquids have been reported by England (Ref. 2) for a small experimental furnace. The results shown in Figure 2 show that the NO_x emissions correlate well with total fuel nitrogen content. He has also shown that emissions under staged conditions correlate with overall fuel nitrogen content. The results suggest that nitrogen containing liquid fuels do not exhibit the wide differences in fuel nitrogen evolution seen in coal.

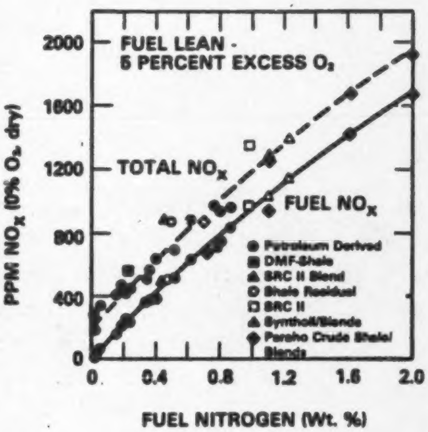

Figure 2. The effect of fuel nitrogen content on total and fuel NO_x (Ref. 2)

Combustion Parameters

The key to the success of advanced concepts is the control of combustion parameters such as air/fuel mixing, fuel-rich residence time, and/or rate and location of heat extraction. The small scale studies discussed above and larger scale experiments have provided valuable insights on the approaches that can be taken to achieve optimum combustion conditions for minimum NO_x emissions. Three techniques that form the basis for the advanced concepts to be discussed are burner design, prechamber burner, and novel concepts.

Burner design. The advanced concepts that are most readily adaptable to existing furnace design are based on controlling air/fuel mixing aerodynamically by changes in burner design. This approach involves fuel and air introduction schemes which create a fuel-rich core in the flame for fuel NO_x control and a surrounding air-rich zone for carbon burnout. The distributed mixing coal burner is an example of this approach.

Prechamber burner. Considerably better control of combustion parameters can be achieved by providing a physical separation of the fuel-rich zone and the subsequent burnout zone. In this approach the thermal environment can also be better controlled. A prechamber burner provides these conditions, and minimum emissions can be achieved as described for the heavy oil burner.

Novel concepts. For certain applications it is possible to abandon the diffusion flame entirely and use a new approach to fuel combustion. This is particularly true for gaseous fuels where mixing problems are minimal and control of thermal environment is a key parameter. Catalytic combustion is an approach described below.

RECENT DEVELOPMENTS

The EPA has an ongoing program for the development of advanced combustion modification techniques for fossil and alternative fuels. This section will describe the status of selected advanced concepts including: 1) distributed mixing systems for pulverized coal fired boilers; 2) sorbent injection through multistage burners for SO_x control; 3) fuel and air distribution techniques for spreader stoker boilers; 4) prechamber furnaces for liquid fuels; and 5) catalytic combustors for gaseous fuels.

Distributed Mixing Systems

The distributed mixing system involves changes in the air and fuel introduction hardware of pulverized coal fired boilers to arhieve aerodynamically staged combustion. The technique has been developed for both wall-fired and tangentially fired boilers and is being evaluated at field installations.

Wall-fired boilers. The system for wall-fired boilers is called the distributed mixing burner (DMB). The DMB concept was developed from EPA sponsored work at the International Flame Research Foundation and has subsequently been

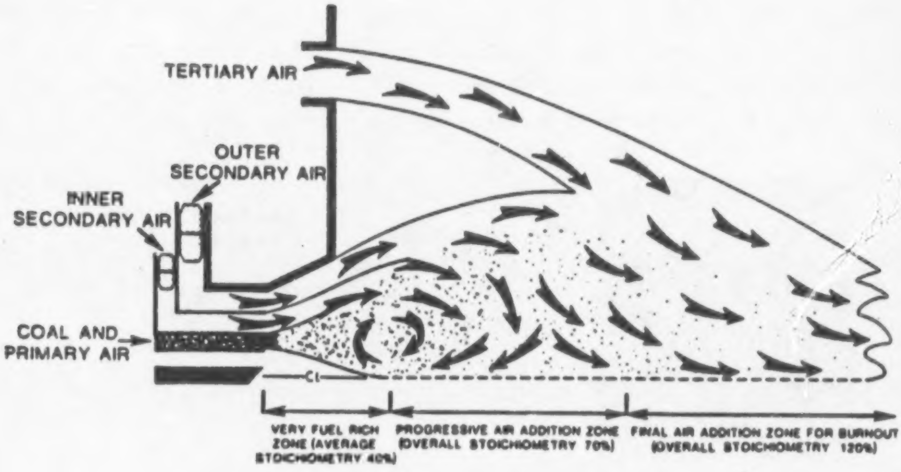
TERTIARY AIR
OUTER SECONDARY AIR
INNER SECONDARY AIR
COAL AND PRIMARY AIR
CL
VERY FUEL RICH ZONE (AVERAGE STOICHIOMETRY 40%)
PROGRESSIVE AIR ADDITION ZONE (OVERALL STOICHIOMETRY 70%)
FINAL AIR ADDITION ZONE FOR BURNOUT (OVERALL STOICHIOMETRY 120%)

Figure 3. Schematic of distributed mixing burner concept

scaled up to full scale for practical boilers (i.e., 30 thermal kilowatt or over 100 x 10^6 Btu/hr). The basis of the burner is shown in Figure 3. The coal is introduced through a central fuel injector with about 25 percent of the combustion air (called primary air). Secondary air is introduced through concentric passages of the burner throat in an amount to bring the total air through the burner to 60 to 80 percent of the air theoretically required for complete combustion of the coal. The initial coal devolatilization and reaction takes place in this fuel-rich zone, thereby minimizing fuel nitrogen conversion to NO_x. Finally the balance of the combustion air, totaling 120 percent theoretical air, is introduced through tertiary ports surrounding the burner. This gradually mixes with the rich primary and provides for complete carbon burnout. An artist's conception of the burner installed in a boiler is shown in Figure 4. Large pilot scale testing in an experimental system has shown the potential for NO_x emissions below 86 nanogram per joule (0.2 lb/10^6 Btu or about 150 ppm).

The DMB concept has been incorporated into a prototype burner (24 thermal megawatt or 80 x 10^6 Btu/hr input) and tested in the experimental furnace. The emission performance is shown in Figure 5 (Ref. 3) for two fuels. For one coal, an Indiana bituminous which is used at the field installation, the effect of secondary air parameters is also shown. At the optimum condition, both coals give NO_x emissions below 129 nanogram per joule (0.3 lb/10^6 Btu). In all cases CO emissions were below 100 ppm.

The prototype burner design has been installed in a four burner, 20 electrical megawatt boiler at the Western Illinois Power Cooperative's Pearl station. The burner has been operating in a nominally staged mode for 6 months, and full staged operation will commence in the summer of 1982. Preliminary estimates indicate that NO_x emissions below 86 nanogram per joule can be achieved.

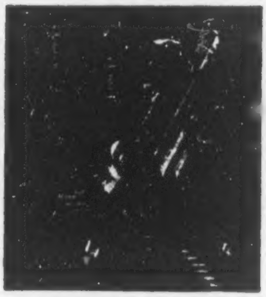

Figure 4. Artist's conception of DMB installed on a boiler

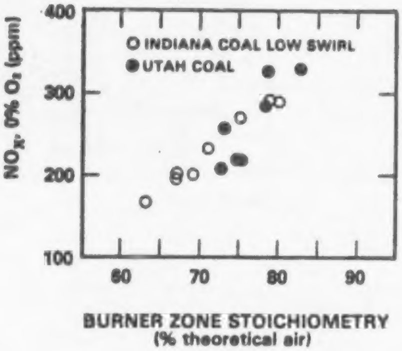

Figure 5. Prototype burner performance (Ref. 3)

Tangentially fired boilers. Since tangentially fired boilers are furnace-stabilized systems rather than burner-stabilized systems, a different embodiment of the distributed mixing approach is required. A schematic of the rich fireball system is shown in Figure 6. The coal and primary air are fired on the tangent of an imaginary circle as in conventional practice; however, the balance of the combustion air is diverted closer to the furnace wall. This produces a fuel-rich zone in the center of the fireball for NO_x reduction, and an air-rich zone for carbon burnout and protection of the furnace wall from corrosion. This system has been tested at small pilot scale (Ref. 4) and showed the potential for NO_x emissions below 129 nanogram per joule.

The rich fireball has been installed on a 400 electrical megawatt boiler at the Hunter station of Utah Power and Light Company. In recent shakedown tests NO_x emissions approaching 129 nanogram per joule have been achieved. Once the operating conditions for continuous operation are selected, the system performance will be evaluated over an 18 month period.

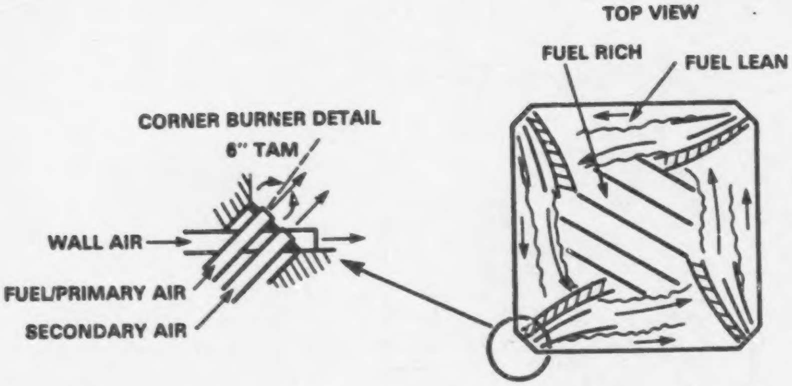

Figure 6. Schematic of rich fireball for tangentially fired boilers (Ref. 4)

Limestone Injection Through Multistage Burners (LIMB)

An analysis of theoretical data suggested that the DMB concept might create conditions suitable for sulfur capture by sorbent materials. Zallen (Ref. 5) ran a preliminary series of experiments with a 3 thermal megawatt (10 x 10^6 Btu/hr) DMB in an experimental facility and reported high capture efficiencies (i.e., up to 70 percent SO_x reduction with a calcium to sulfur ratio of 2). Various reasons for the high capture were postulated, including: 1) reduced peak temperatures, 2) existence of a fuel-rich zone where reduced sulfur species are captured directly, 3) intimate sorbent sulfur contact, and 4) in-situ activation of the sorbent. As a result of these findings the EPA initiated a development program to confirm the potential of the technique to establish the critical parameters for sulfur capture and to define approaches to practical application. To date the effort has been concentrated on bench scale experiments and prototype burner testing.

Bench scale results. The bench scale studies have been carried out by Case (Ref. 6) in a 300 thermal kilowatt (1 x 10^6 Btu/hr) facility called the Boiler Simulator Furnace (BSF). The BSF is designed to provide control of temperature and/or residence time of combustion products from the burner zone to the stack over the range typical of practical boilers. The furnace variables include, burner zone stoichiometry, sorbent injection, heat extraction, and fuel type. Figure 7 compares the range of sulfur capture in the BSF with past data from experimental systems and for furnace injection of sorbent for a practical boiler. The broad crosshatched area, which shows the BSF data, indicates that sulfur captures can range from 10 to 70 percent at a calcium to sulfur ratio of 2, depending on operating conditions. For overall fuel-lean combustion zone operation, the primary variable is temperature, with high captures achieved with high heat extraction from the radiant zone. For staged operation the capture appears to be less sensitive to temperature. Further work is in progress to fully identify the controlling parameters.

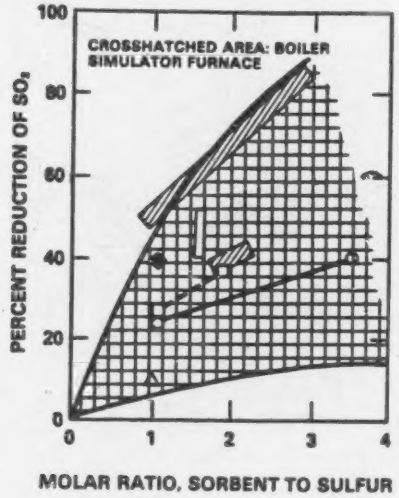

Figure 7. Range of sulfur capture achievable in BSF compared to data for other systems

Prototype testing. The DMB prototype burner discussed earlier has been evaluated for sulfur capture by sorbent injection in a limited series of tests in two large experimental facilities (Ref. 6). One facility, the large watertube simulator (LWS), has a thermal environment similar to industrial coal fired boilers. The other, the medium tunnel (MT), has a hotter burner zone environment. The sorbent was either pulverized with the coal or injected with the tertiary air when the burner was operating at low NO_x conditions. Data for prototype testing is shown in Figure 8a for the LWS for two coals. For the high sulfur Indiana coal, the sulfur capture was essentially the same regardless of sorbent injection location and was about 50 percent at a calcium to sulfur ratio of 2. For the low sulfur Utah coal the sulfur capture was lower than for the Indiana coal, and injection with the coal gave better results. Data from the MT in Figure 8b shows similar trends with fuels; however, it appears that in the hotter environment, sorbent injection with the coal provides better capture than injection with the air.

Practical application. Application of the technology will be dependent on the sulfur capture potential achievable in the combustion environment of a range of practical boilers. In addition the impacts of the sorbent addition on boiler operation (e.g., slagging and fouling) and on post treatment device efficiency (e.g., particulate control equipment) must be evaluated. An engineering design study is in progress to address these and other questions.

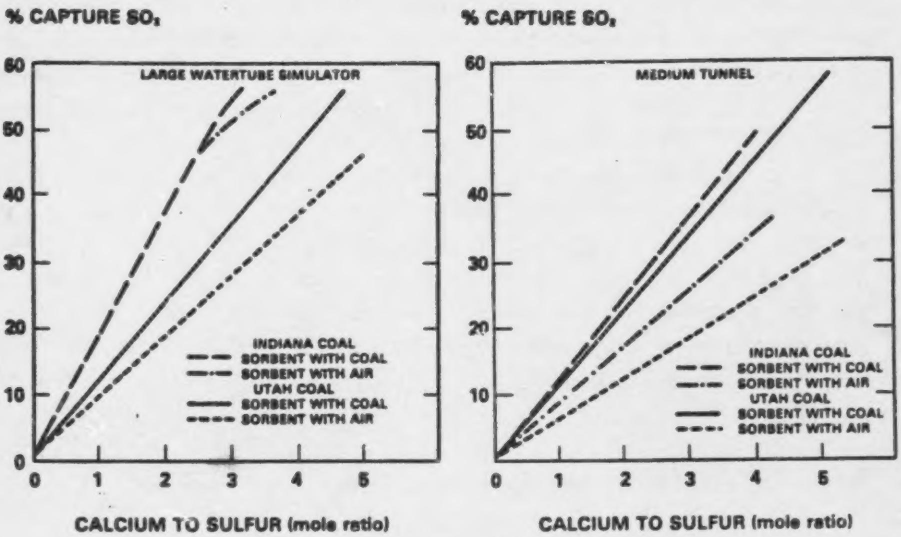

Figure 8. Sulfur capture with prototype burner in: a) the large watertube simulator and b) the medium tunnel

Stoker Coal Fired Concepts

In the past only limited attention has been given to NO_x control for stoker systems although this type of boiler makes a significant contribution to NO_x emissions. Field evaluation of operating changes shows that these changes (e.g., air distribution) are typically much less effective for NO_x control for stokers than for other coal fired boilers. Recent R&D has investigated the combustion processes occurring in stoker systems. In all stoker systems there is a bed of burning coal supported by a grate of some kind, and the phenomenon of bed burning has a significant effect on pollutant emissions. For spreader stokers the coal is deposited on the grate by being "thrown" from a distance over the bed. This results in entrainment of coal fines by hot gases rising from the bed and a portion of the coal fines being burned under conditions similar to those of pulverized coal firing. Both bed burning of the raw coal and suspension-phase combustion of the fines have been investigated in simple experimental systems. Major conclusions are:

1. Even for spreader stokers, between 80 and 90 percent of the coal is burned on the bed. The conversion of fuel nitrogen to NO_x is about 15 percent for fuel-lean beds. The conversion may be reduced to less than 5 percent by operation of the bed in a fuel-rich mode (Ref. 7).
2. Dependent on the amount of fines in the raw coal, between 8 and 20 percent

of the coal will burn in the suspension phase of a spreader stoker. Under fuel-lean conditions about 40 percent of the fuel nitrogen will be converted to NO_x. This conversion can be reduced by fuel-rich operation of the suspension zone. It also appears that incomplete combustion of the coal fines can contribute significantly to unburned carbon in the fly ash (Ref. 7).

3. A series of experiments was performed with coal pellets which contained limestone sufficient to give a calcium to sulfur ratio of 3.5. For fuel-lean operation the SO_2 emissions are reduced from 3655 to 1290 nanogram per joule (8.5 to 3.0 lb $SO_2/10^6$ Btu). Under staged combustion conditions the SO_x is further reduced to about 645 nanogram per joule (1.5 lb $SO_2/10^6$ Btu). This shows the potential importance of fuel-rich conditions on sulfur capture. This effect needs to be explored for lower calcium to sulfur ratios to determine if better enhancement is possible (Ref. 8).

As a result of these investigations, several control concepts have been postulated and tested (Ref. 7) in a model spreader stoker as shown in Figure 9. In this system coal is fed to the spreader which distributes the coal onto a fixed grate. Combustion air is supplied from under the grate and through overfire air ports to simulate arrangement in practical systems. The results obtained include:

1. For staged combustion the amount of air introduced under the bed is reduced and the amount of overfire air is increased to maintain a constant overall level of 1.4 times the amount required for complete combustion (SR = 1.4). As the bed stoichiometric ratio (S.R.) is reduced, NO_x levels decrease and a maximum reduction is achieved at an S.R. of 0.8. The level of reduction depends on the height of overfire air injection, with the larger separation giving lower NO_x as shown in Figure 10. At the optimum conditions a 65 percent reduction was achieved.
2. Since the fines combustion accounts for a significant portion of the NO_x from spreader stokers, an experiment was run where the fines (i.e., particles smaller than 0.279 cm or 0.111 in.) were removed from the coal feed prior to combustion. This relatively simple change gave a 20 to 25 percent reduction of NO_x. Staged combustion further reduced the emissions; however, overfire air height had relatively little effect, indicating the significance of suspension zone reactions on NO_x formation (Ref. 9).

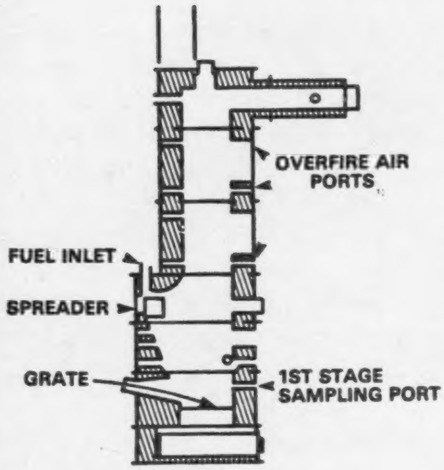

Figure 9. Schematic of model spreader stoker (Ref. 7)

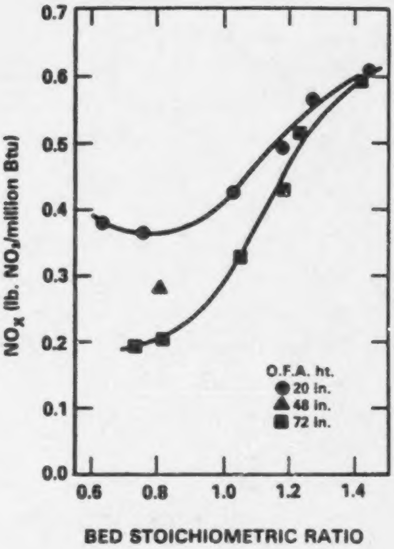

Figure 10. Effect of bed stoichiometry and overfire air height on NO_x from model spreader stoker*

Liquid Fuel Burner

The small scale fuel screening of petroleum and alternative liquid fuels discussed earlier also led to a concept for highly effective control of fuel NO_x. Staged combustion experiments showed that, if the heat loss from the fuel-rich zone was very low, NO_x emissions could be reduced over 90 percent for a 2 percent nitrogen shale-derived crude. It further showed that under optimum staged conditions NO_x levels were relatively independent of the nitrogen content of the oil. Based on these results a prechamber liquid fuel burner concept was derived for application to package boilers and process heaters (Ref. 10). A 600 thermal kilowatt (2×10^6 Btu/hr) prototype burner, as shown in Figure 11, was constructed and tested in an experimental facility similar to a scotch marine firetube boiler. The performance of the prototype burner with a 0.98 percent nitrogen coal-derived liquid fuel is shown in Figure 12. At a burner zone SR of 0.7 (70 percent of the air required for complete fuel combustion) NO_x emissions were reduced to 55 ppm with a very low smoke number. This represents a reduction of over 90 percent from the base level of 850 ppm. The performance of the prototype for a range of high nitrogen fuels was evaluated, and the NO_x emissions for the prototype burner were below 80 ppm for all liquid fuels. Also the prototype produced lower NO_x emissions than the screening furnace, probably due to surface to volume effects which reduce heat loss at the larger scale. The prototype burner provided the following design guidance: 1) the first stage should be run as hot as possible, 2) the fuel-rich residence time should be greater than 200 ms, and 3) second-stage heat removal and air injection are not primary variables.

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*NOTE. It is EPA policy to use metric units; however, for figures derived from references using English units the units are not converted to metric. For metric equivalents see table at the end of the paper.

Based on these results a 3.6 thermal megawatt (12 x 10^6 Btu/hr) prototype has been tested and produced similar NO_x emissions performance. The burner will subsequently be scaled up to 18 thermal megawatt (60 x 10^6 Btu/hr) and tested on an oil field steam generator.

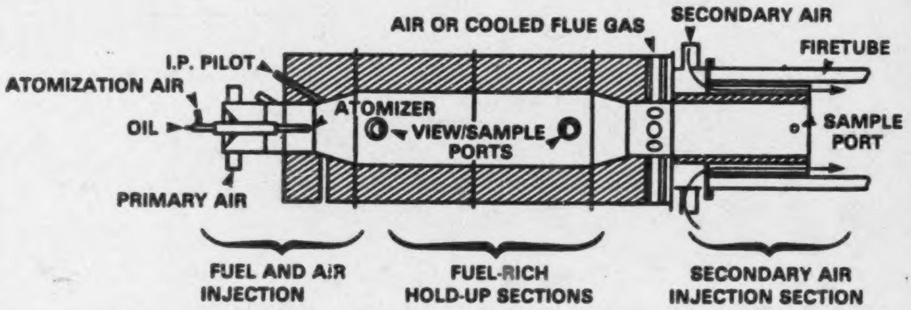
AIR OR COOLED FLUE GAS
SECONDARY AIR
FIRETUBE
I.P. PILOT
ATOMIZATION AIR
OIL
ATOMIZER
VIEW/SAMPLE PORTS
SAMPLE PORT
PRIMARY AIR
FUEL AND AIR INJECTION
FUEL-RICH HOLD-UP SECTIONS
SECONDARY AIR INJECTION SECTION

Figure 11. Schematic of heavy liquid fuel burner (Ref. 10)

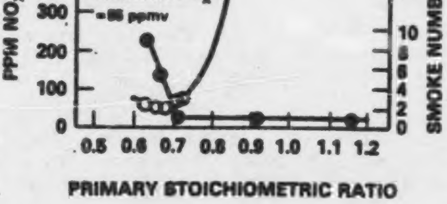

Figure 12. NO_x emissions for optimized heavy liquid fuel burner

Catalytic Combustion

The use of catalysts as cleanup devices has been known for some time; however catalysts have been explored as a primary combustion device only in the last few years. In this application the catalyst is used to promote the combustion of the fuel and serves as a major energy release device in a combustion system. The catalyst shortens reaction times and can result in very low levels of pollutants (NO_x, CO, and unburned hydrocarbon) if properly applied. The main problem is interfacing the catalyst with the practical combustion system. A number of applications have been studied (Ref. 11). To date the most success has been achieved with gaseous and light liquid fuels, since the catalyst typically requires a homogeneous air/fuel mixture, with emphasis on control of thermal NO_x.

One of the most advanced concepts is the fiber burner for application to gas fired package boilers and process furnaces. A high temperature ceramic fiber is deposited over an external support screen. Premixed air and fuel are fed to the inside of the fiber burner, pass through the fiber cylinder, and are ignited and burn on the outer surface. The surface is placed in close proximity to a heat transfer surface so that about 50 percent of the energy release is transferred from the burner surface by radiation. This serves two purposes: 1) it maintains the temperature of the ceramic fiber well within its operating limit, and 2) it reduces the formation of thermal NO_x. A schematic of a fiber furnace for a scotch marine firetube boiler is shown in Figure 13 (Ref. 11). A prototype system has been constructed and tested in a 300 thermal kilowatt (10^6 Btu/hr) firetube boiler. The NO_x emissions are shown in Figure 14a and boiler efficiency is shown in Figure 14b. At 100 percent load firing natural gas, NO_x emissions are 10 ppm as compared to 70 ppm from the conventional burner. At this condition CO and unburned hydrocarbon (UHC) emissions are less than or equal to those of the conventional burner. It also appears that a portion of the NO_x, CO, and UHC emissions are attributable to thermal combustion of the fuel/air mixture short-circuiting through fiber pad imperfections. The boiler efficiency is increased by

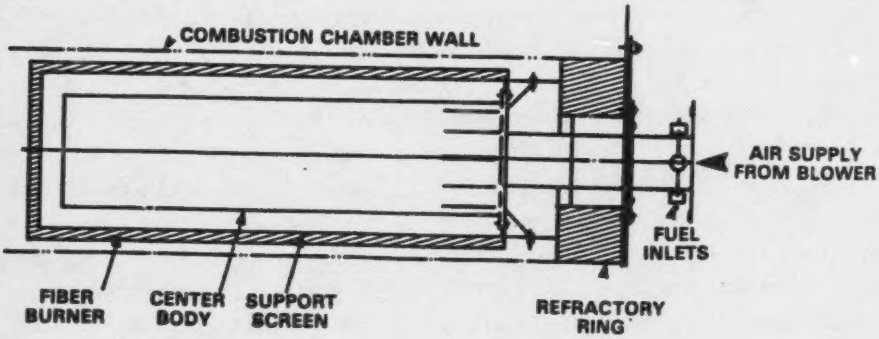

Figure 13. Schematic of fiber burner assembly (Ref. 11)

about 1 percent at 100 percent load and normal excess air loads. Since the fiber burner can operate at lower excess air and higher load without excessive CO and UHC emissions, the system efficiency can be increased an additional 0.5 percent at 120 percent of rated steam production.

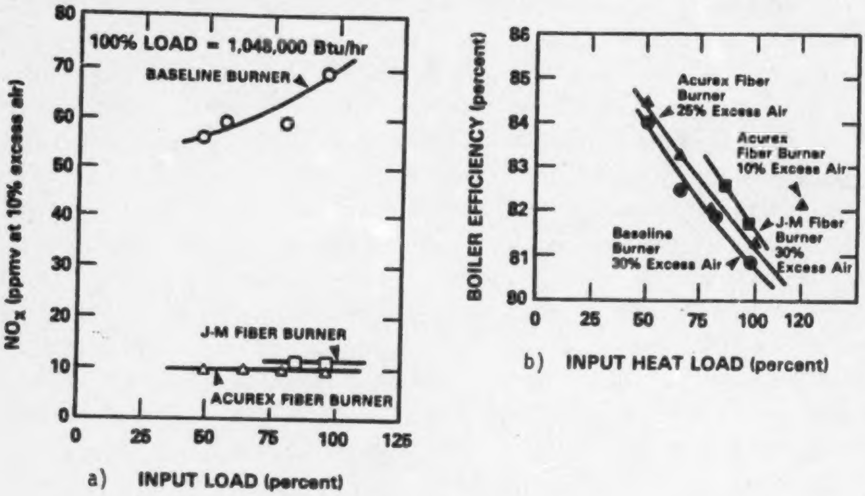

Figure 14. Performance of fiber burner in a firetube boiler:
a) NO_x emissions; and b) efficiency

SUMMARY

The EPA program for advanced combustion modification technique development has provided an understanding of the effects of combustion conditions and fuel characteristics on NO_x formation and control. The main focus of the program has been controlling fuel NO_x emissions from nitrogen containing solid and liquid fuels. Fundamental and small pilot scale work has provided insight into the role of fuel nitrogen volatility on NO_x formation from coal, heavy oil, and synthetic liquid fuels.

The advanced combustion modification development program has placed the greatest emphasis on staged burner development for control of fuel NO_x from high nitrogen solid and liquid fuels. In addition, novel combustion techniques for clean fuels have been investigated. A summary of the developments described in this paper follow.

1. Prototype testing of distributed mixing systems for pulverized coal firing has shown the potential for controlling NO_x emissions to the level of 86 to 129 nanogram per joule (0.2 to 0.3 lb/10^6 Btu). Field evaluation results from retrofit of these systems indicate that emissions will fall at the upper end of that range. New systems specifically designed for these firing systems appear to have the potential to achieve the lower end of the range.

2. Several novel systems for stoker coal have been developed based on fuel rich operation of the bed region and control of conversion of fuel nitrogen in the small coal particles. Further development is required; however, it appears that NO_x reduction over 65 percent is possible.

3. A preburner system for high-nitrogen liquid fuels has been scaled up from 21 thermal kilowatts to 3.6 thermal megawatts (70,000 to 12 x 10^6 Btu/hr). NO_x emissions from a given fuel generally decrease with increased scale. Absolute NO_x levels are below 100 ppm with very low carbonaceous particulate emissions.

4. The fiber burner approach to catalytic combustion appears to be feasible for package boilers and process furnaces. Evaluation in a small firetube boiler indicates a potential for NO_x reductions up to 85 percent with low levels of CO and unburned hydrocarbons. Both steam generation capacity and system efficiency are increased.

5. The use of sorbents in conjunction with low NO_x combustion techniques for pulverized- and stoker-coal-fired systems appears to have potential for substantial SO_x control. Based on large prototype burner testing and bench scale experiments, LIMB has the potential to achieve 50 percent SO_x reduction on retrofit systems and a higher degree of control for new systems.

6. These and other advanced combustion modification techniques under development have the potential for highly effective NO_x control at moderate cost. The coupling of low NO_x and SO_x is a particularly attractive technical option, assuming it can be accomplished at reasonable cost.

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Metric Equivalents

0.1 lb/10^6 Btu = 43 nanogram/Joule
1 inch = 2.54 cm
10^6 Btu = 300 thermal megawatts

This Babcock and Wilcox (B&W) report, which was presented at the Joint Symposium on Stationary Combustion NOx Control, updates information on dual register burner[illegible]issions of NOx from fossil-fueled power plants using B&W dual regis[illegible]rs has been reduced by up to 60 percent at modest cost.

Technical Paper

Fossil Steam Generator NOx Control Update

J. A. Barsin, Member ASME
Manager, Engineered Products
Babcock & Wilcox
Barberton, Ohio

Presented to
Joint Symposium on Stationary
Combustion NO_x Control
Dallas, TX
November 1-4, 1982

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Fossil Steam Generator NO_x Control Update

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PGTP 82-110

Since the third NO_x Control Seminar of 1980, additional information has been gathered concerning actual NO_x emissions from fossil-fuel power plants equipped with B&W Dual Register Burners/Compartmented Windbox furnace systems. These systems use a wide range of both bituminous and subbituminous coals. The field results, from actual units firing coal and oil, have demonstrated NO_x reduction of up to 60 percent compared to units without NO_x control.

B&W's future low-NO_x R&D direction is summarized as well as the laboratory work done on the High Nitrogen Liquid Fuel burner (HNLF). Also, the most recent laboratory-facility additions (a new $5x10^6$ Btu/hr furnace, atomization facility and laboratory ashing furnace) are described. These additions will permit more extensive small-scale work. Retrofit concepts for the pre-NSPS coal-fired units, both cyclone and pulverized coal, are presented. Furthermore, a summary is given of B&W's future low-NO_x R&D direction for meeting the increasingly restrictive NO_x-emission levels.

The Dual Register Burner/Compartmented Windbox System has been commercially applied to all New Source Performance Standard (NSPS) steam generators sold domestically by B&W since 1972. At the last NO_x Symposium (1980), it was reported that 26 steam generators representing 15,000 MW of pulverized coal-fired capacity had been placed in service utilizing this system. Now in 1982, the numbers are 34 units and 20,000 MW. This represents 2,000 burners in service and 3,500 shipped in ten years, firing bituminous, subbituminous and lignite fuels (Figure 1). Detailed descriptions covering the laboratory development and initial retrofit of this system can be found in References 1 and 2.

The largest unit to come on steam during the past

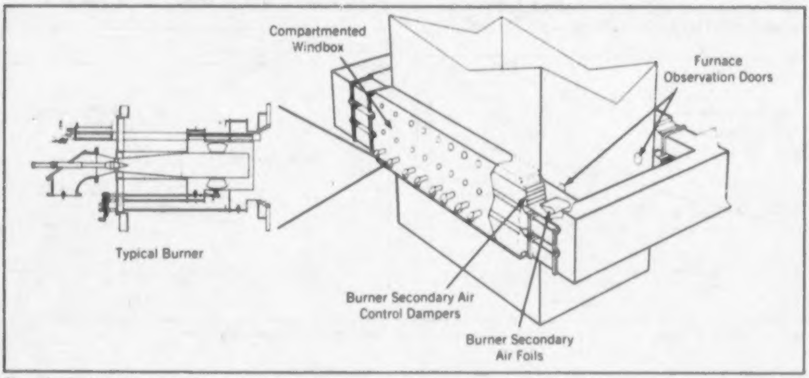
Compartmented Windbox
Furnace Observation Doors
Typical Burner
Burner Secondary Air Control Dampers
Burner Secondary Air Foils

Figure 1 Dual register burner/Compartmented windbox system

two years was AEP's Mountaineer No. 1, a 1300-MW unit with two more 1300's to follow at the Rockport Station. Typical additions to the B&W data base were in the 600-MW size range; good examples of which are Basin Electric's Laramie River Nos. 2 and 3, SWEPCo's Welsh Nos. 2 and 3 and Louisville Gas & Electric, Mill Creek No. 4.

All of our steam generator NO_X guarantees were based upon the data generated from the Dual Register Retrofit at the E. C. Gaston Plant, Southern Company's Southern Electric Generating Co.[3] The NO_X prediction curves developed by that data, correlating heat input per adjusted furnace surface (HA/Sc,) vs NO_X expected for various excess air levels with baseline circular burner, followed up by Dual Register Burner performance, permitted us to offer low NO_X systems that could meet the NSPS[2]. Plotting recent field data on the 15 percent excess air curve generated from that retrofit (Figure 2) indicates that, in retrospect, we were conservative and our risk in making the guarantees was minor. The NSPS was revised in 1979 from 0.7 lbs $NO_X/10^6$ Btu for all coals, except lignite, to 0.6 for bituminous and 0.5 for subbituminous coals. The B&W low NO_X system, though designed for the 1970 NSPS, meets or betters the 1979 revised standards without having to go to substoichiometric conditions (two stage) in the furnace. Our system does not require "boundary air" or diversion air/overfire air to meet the NSPS of 1979.

The first Dual Register Burner went into service in 1975. The seven years of operating experience has allowed us to evaluate the system's wear life and mechanical operation. This review has resulted in several revisions to the base burner which include the following: 1) The venture/plug combination has been removed because of marginal coal distribution, severe erosion and high pressure drop. It has been replaced with the conical diffuser, which is cast of wear resistant material, has a lower pressure drop, excellent peripheral coal distribution, and weighs less than 200 pounds. 2) The inner air register has been eliminated because the tangential components of air flow that it produced were subsequently destroyed by the swirl vanes. That register has been replaced with a sliding sleeve to control air flow quantities in that zone. 3) The spin vane operational gearing did give movement problems initially due to fly-ash packing in the gears. A redesigned gear was retrofitted on jobs requiring it and now is the standard. 4) Recently, several coal nozzles on some jobs have been overheating when out of service on upper level burners. The cause is related to out-of-service compartment cooling air flows lower than recommended. The stainless 12-inch long nozzle was not damaged but some of the carbon steel coal pipe located behind the stainless nozzle has demonstrated overheat, and the alloy zone will be extended an additional 6 inches for a total of 18 inches to correct this problem.

Laboratory work on new low NO_X concepts in progress during the past two years has concentrated on developing a short flame low NO_X burner for the smaller industrial coal-fired units and a good low NO_X burner suitable for coal, gas and oil firing. The low NO_X P.C. burner suitable for gas, oil and coal fuels is a joint development under funding and direction of the Stork Boiler Company of the Netherlands. The burner is similar to that developed jointly with Deutsche-Babcock in 1978, at a five-million Btu level (Figure 3). These trials are at the 30×10^6 Btu level and provide us with another laboratory data point to reduce risks associated with the commercial application. This laboratory-developed burner will be applied to a 600-MW unit currently under design.

Our Industrial and Marine Division continues its development of a short flame low NO_X burner

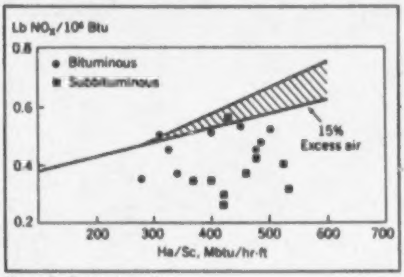

Figure 2 Field test results

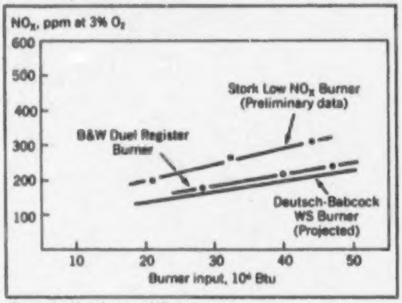

Figure 3 Multifuel low NO_X burners

suitable for use in small industrial furnaces. The test furnace utilized for these laboratory trials was a FM (Factory Assembled) package boiler rated at 50,000 lbs/steam per hour on oil or gas. It was not designed for coal but performed satisfactorily from a slagging, fouling and carbon utilization standpoint at 60 percent MCR for about 45 days of testing. Three different concepts were tried, all providing near field mixing flexibility and were compared to the two baseline conditions of circular burner (high NO_X) and standard Dual Register (low NO_X). The first trial utilized a revised standard circular burner impeller which we labeled "Radial Vane Impeller". The second concept labeled "Adjustable Swirl" utilized a movable sleeve to vary the tangential components of secondary air flow. The third concept entitled "Controlled Mix" utilized stationary fixed vanes in a dual-air zone configuration. Air flow splits, inner to outer, were controlled by a sliding sleeve. Results (Figure 4) indicate the "Controlled Mix" burner was the lowest NO_X producer and had the shortest flame. Three coals were tested on each concept and the baselines included eastern bituminous, western bituminous and subbituminous. The "Controlled Mix" burner is currently undergoing field evaluations in an industrial retrofit experiment with encouraging initial results.

Our laboratory work on the Low NO_X Combustor System (LNCS) at the 5 and $40x10^6$ Btu levels is complete under joint EPRI/B&W sponsorship. The concept of a fuel-rich (deep staging) first stage, and destruction of NO_X in the second stage, as a function of temperature (Johnson Effect)[4], has been adequately demonstrated in the laboratory and awaits adequate funding to cover a field demonstration. The cost analysis made in 1980 continues to indicate that combustion modifications are the most cost effective way to go. Selective Catalytic Reduction (SCR) cost estimates lie between 55 and 100 $/kW capital and 1.7 mils/kWh operational. Thermal deNO$_X$, although at 10 $/kW is a capital bargain compared to SCR, has high operational costs of 1.2 mils/kW. The licensor of thermal deNO$_X$ indicates that a 30 percent reduction on coal, due to limited coal experience obtained to date, is a level they could guarantee. Capital costs associated with the B&W/EPRI LNCS (a combustion modification) are less than $15/kW for an 80 percent reduction and operational costs are less than 10 percent of the thermal deNO$_X$.

B&W low NO_X coal development goals are summarized in Figure 5. Systems providing up to 80 percent reductions from uncontrolled levels have been described. The system labeled "After-Burning," coupled with LNCS, has the potential to effect 90 percent NO_X generation reductions when compared to uncontrolled levels. A numerical goal attached to that percent reduction would be the achievement of less than 100 ppm of NO_X on a P.C. system (Figure 6).

Laboratory work reported by Babcock Hitachi and others on the IFNR (In Furnace NO_X Reduction) or after-burning approach would couple the flexibility of our present compartmented windbox system with the LNCS. The compartmented windbox permits individual pulverizer air control that, in turn, permits the lower burners to operate at a 15 to 80 percent stoichiometry, providing an overall lower furnace first-stage stoichiometry of 0.7 percent. This level has already been laboratory demonstrated for low NO_X potential by the LNCS work, and now, by many other investigators. Several of our NSPS steam generators, equipped with compartmented windboxes and OFA ports, could be utilized to field demonstrate the LNCS/after-burning low NO_X potential. The after-burning fuel injectors could use traditional primary air coal ratios or dense-phase transport to maintain minimum O_2 levels. Both approaches should be investigated. The delay in scheduling a field demonstration is due to the prob-

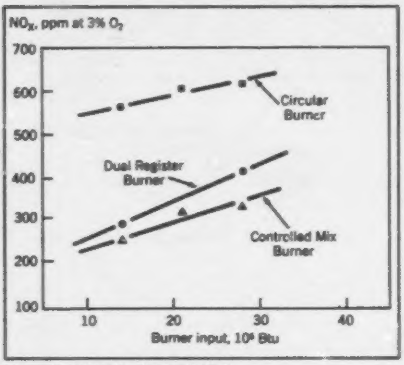

Figure 4 B&W FM boiler (coal firing)

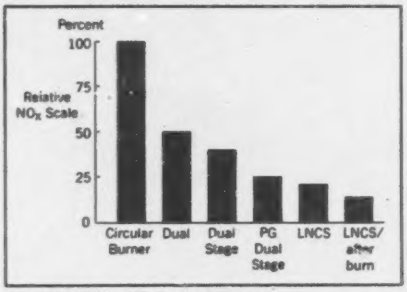

Figure 5 B&W low NO_X (coal) development

lem of obtaining financial assistance.

Fuel nitrogen correlations from the extensive field data collected have not correlated well with the NO_x emissions measured. Our Research Division had started to develop a fuel nitrogen/temperature correlation, but the data developed by Pershing under IEA funding, correlating the fuel N, of 48 worldwide coals with emissions, appeared to be so promising that we will redirect our efforts and attempt to verify those stage/excess air correletions, and test our field experience against the laboratory generated data.[5]

Cyclones

Cyclone furnaces were designed to burn midwestern bituminous coal, high in iron and sulfur, having medium ash fusion temperatures. The commercial application was successful and a total of 1,100 cyclone furnaces presently exist. The cyclone, in order to work properly, requires coal crushed to the correct size, a slag layer coating the inner surface of the barrel to hold the coal particles, and a slag tapping capability that has demonstrated up to 75 percent slag removal from the flue gas stream. Local clean air regulations have encouraged the shifting of fuel from high-sulfur original design coals to low-sulfur western fuels. The ash fusions associated with the western replacement subbituminous coals are low and do not permit the development of a barrel slag layer. This lack of slag in the barrel leads to increasing unburnt carbon carryover. We have established a development program to modify the present cyclone design to better handle these off-design subbituminous coals and that program is underway. Additionally, it has become apparent that a fundamental study in NO_x reduction should be undertaken. The unstaged cyclone furnace, firing a midwest bituminous coal can and does, when adjusted properly, generate NO_x levels in excess of 1200 ppm. Concepts emerging from the various MHD programs indicate "superhot" is an approach that must be explored and could be applied to the cyclone to reduce NO_x. The classical two-stage approach is not appropriate with the base design fuels due to the potential for catastrophic corrosion. Various novel concepts advanced in house and by John Pohl of EER and Steve Lanier of U.S. EPA, coupled with the commercial availability of bimetallic corrosion resistant tubing, now offer the potential for the practical application of two-stage fuel rich combustion (Figure 7). The direct firing of coal water slurries in a cyclone furnace (1960), using 65 percent coal/35 percent water, proved successful at that time and as a side effect aided in controlling railroad coal rates by adding an alternative transport mode. Highly loaded slurries (70 percent coal plus) are now a potential fuel source that could, in the cases studied, lower NO_x significantly in a cyclone due to the reduced flame temperatures and the reduction in the thermal NO_x contribution. The coal-fired cyclone, in a modified form to achieve 90 percent ash removal, could be a clean gas generator for steam generators presently firing natural gas or oil. The cyclone developmental program now underway at B&W includes updated designs for the current cyclone population to better handle actual vs design fuels, the work on the potential clean gas generator for retrofits, and the superhot low NO_x generator.

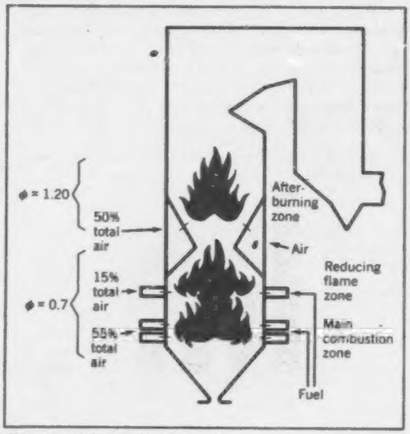

Figure 6 LNCS/after-burn concept

Retrofits

Ambient air quality degradation and proposed acid rain legislation, where NO_x reductions can be

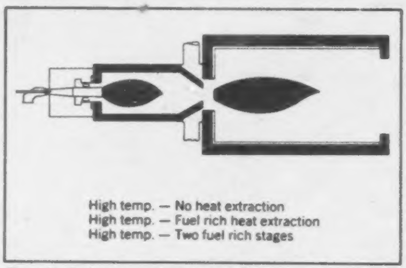
High temp. — No heat extraction
High temp. — Fuel rich heat extraction
High temp. — Two fuel rich stages

Figure 7 Cyclone concepts to be examined

credited against SO_x required reductions, indicate that existing old stationary sources may be candidates for low NO_x retrofits. Described already are the concepts planned for application to the cyclone furnaces; what follows is a description of our pulverized coal program.

The Distributed Mixing Burner (DMB) EPA concept developed on the laboratory scale by EER, Inc., and applied to a small utility boiler by EER/FW does, under staged conditions, offer an alternative low NO_x device for new steam generators. Our project, under EPA funding, was to locate a large utility host site to retrofit the B&W version of the DM burner. The clients who were agreeable to such a demonstration did not possess host sites that met all the site-selection criteria and we were not able to perform the retrofit evaluation. This disappointing outcome resulted in a redirection of the program. It will now consist of the field evaluation of a circular burner and Dual Register Burner, to obtain benchmark data. Those burners will then be characterized in the EPA's Large Water Simulator (LWS), along with the B&W version of the DMB. The three burners fired with the same fuel in the same facility will permit us to project actual NO_x levels that would have been attained on a retrofit unit. The results will provide us with information in sufficient detail to permit us to predict commercial guarantee levels similar to what was necessary when only the EC Gaston Dual Register Retrofit Data was available in 1972.

The Dual Register Burner, when developed, had to be installed as a retrofit because we could not afford to wait the typical 5-year span associated with date of sale/commercial operation, typical of the early 1970s. In the case of that host site, S.E.G.Co., E.C. Gaston Unit No. 1, the retrofit was feasible. The furnace size, the existing burner throats, and the slagging and fouling characteristics of the fuel were all compatible and the retrofit was successful. Each potential retrofit, however, must be studied to insure compatibility with a low NO_x system.

Our series of steam generators sold during the highly competitive 1960s were fitted with turbulent diffusion cell burners and furnaces sized to induce rapid intense combustion. These units are coal fueled and represent approximately 30,000 MW of capacity now approaching the 20-year remaining life point. The typical NO_x generated from these pre-NSPS units ranges up to 900 ppm and provides an opportunity for reductions, if required. Our retrofit concept for the cell burner is a locally-staged low NO_x circular burner. The low NO_x circular burner, provided with swirl control and controlled fuel/air mixing with a conical diffuser, has been single-burner field tested with encouraging results. A full-scale retrofit is desired, but would require external financial support (Figure 8).

Laboratory facilities

Since the last symposium, we have added several major fossil fuel facilities that will permit further investigations into the slagging and fouling characteristics, combustion of new fuels, and lower NO_x concepts. The Laboratory Ashing Furnace (LAF) has been redesigned, rebuilt and returned to service. Micronized fuels have been characterized on this facility and indicate reduced ash deposition rates, but have demonstrated increased compression strengths when compared to the parent pulverized coals. The five-million Btu tunnel has completed the LNCS and micronized firing trials and is currently dedicated to the EPRI/B&W coal water firing trials. Management has authorized the addition of a atomization facility equipped with laser diagnostics to determine particle size, particle distribution and particle velocities. Coal water and limestone atomization studies have been carried out since August 1982, when the facility went into operation. A new five-million Btu furnace, providing wall-firing capability, is under construction and will be operational in February 1983. A solid fuel handling and pulverizing (MPS mill) system has been added to fuel this furnace and permit additional in-house fuel character-

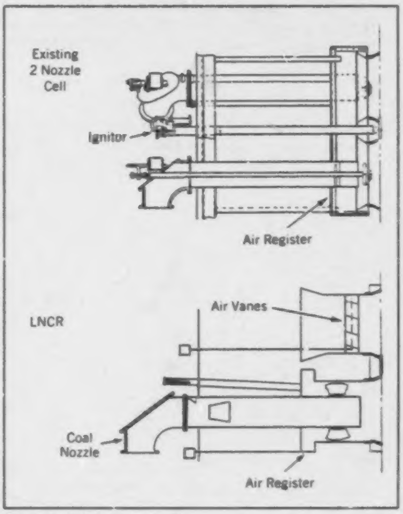

Figure 8 Cell burner retrofit

izations. The LNCS and after-burning concepts will be utilized to aid in the start-up/shakedown/characterization efforts in this facility (Figure 9.)

Oil/gas

The high nitrogen liquid fuel burner (HNLF) has completed 5x10^{6} Btu trials and is now being demonstrated on the Stirling Furnace at 40x10^{6} Btu/hr input. EPRI donated SRC-II with a 1.0% nitrogen content to aid us in evaluating this flexible low NO_X oil/gas burner. No. 6 oil with 0.2 percent nitrogen was also utilized and both fuels demonstrated low NO_X (70 percent less than a circular burner) for this facility. The flexibility of this burner was demonstrated when it was utilized for coal-water firing trials and was successful in obtaining satisfactory turndown, stability, and carbon utilization. This burner represents a retrofit product that could be applied to pre-NSPS units to reduce NO_X.

Researchers have confirmed the effect of atomizer design upon NO_X generation by revising our standard Y-jet drillings slightly to group exit channels. This minor change reduced NO_X by 25 percent using the same base fuel, burner, and furnace. This effect is being field evaluated now and we expect to see potential reductions.

The HNLF burner is projected to reduce NO_X emissions by 75 percent from uncontrolled levels. The HNLF burner coupled to the LNCS/after-burning system is expected to be capable of over 90 percent reductions when compared to uncontrolled sources.

Research and development

Research and Development plans for the 1980s will be concentrated in four major NO_X areas. The first are the fuel nitrogen correlations that govern the NO conversions in both solid and liquid fuels. The second major area is to obtain field demonstrations of the LNCS/after-burn system, cell burner retrofit, and HNLF burner retrofit. The third area is to continue development work on the cyclone furnace concepts and apply them to commercial-sized retrofits once laboratory results are complete. The fourth area concerns synthetic fuels and our continued work characterizing and burning such fuels as SRC I and II, COM, CWS, Exxon Donner Solvent product fuel, char and other liquefaction products from coals. We have gained extensive experience with some of these fuels under DOE, EPRI and in-house funding; however, the burning, fouling, slagging and handling characteristics of some of these fuels require further work.

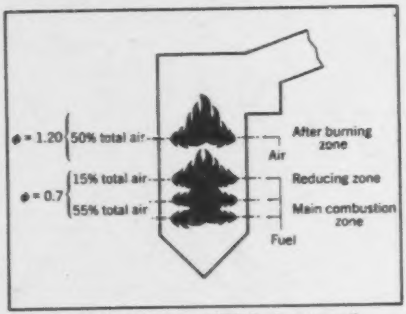

Figure 9 5MMbtu test furnace (LNCS/after-burn concept)

Summary

Results from the 34 pulverized-coal units in service to date show that the Dual Register Burner/Compartmented Windbox System is an effective tool for NO_X control. NO_X levels up to 60 percent lower than those achievable with the high-turbulence circular burner system are obtained on approximately 20,000-MW NSPS units. Additional field demonstrations applying the LNCS/after-burn system, cell burner retrofit and cyclone furnace updates are required to meet the increasingly strict NO_X emission limitations expected during the next six years. This applies to both new sources and old sources.

The combined effects of industrial recession and conservation have restricted traditional utility growth rates but it is anticipated that by 1985, we will have applied the Low NO_X Dual Register System to 125 utility steam generators with an approximate capacity of 60,000 MW. Typically, these units will generate half of the NO_X they would have had if the New Source Performance Standards of 1971 had not implemented emission controls.

Acknowledgments

The author appreciates the assistance of T. Sommer, Combustion Technology; A. LaRue, Combustion Specialist; and J. Fairbanks, NO_X Specialist of Fossil Power Systems; G. Lindstrom, G. Farthing, T. Morris and C. Eckert of the Chemical Engineering Section, R&D Division; and C. Lane, Manager, Fuel Equipment Design, Industrial Power Generation Division for their help in obtaining and reducing the data upon which this report is based.

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Burner design has a significant effect on the formation of NOx gases. Dual register burners reduce flame temperature and regulate oxygen availability, thus reducing NOx formation. Application to both oil and coal-burning boilers is discussed in this Electric Power Research Institute (EPRI) report.

ADVANCED BURNER/FURNACE DESIGNS

Combustion research studies have shown that burner and furnace configurations have an important influence on NO_x formation. Since these designs involve the control of very complex processes of fuel and air mixing, ignition, heat transfer, flow re-entrainment and interactions, it is not surprising that the development of low NO_x combustion systems has been a slow, arduous process. In the interim, utilities faced with NO_x compliance regulations have taken the most expedient approach by applying combustion modifications using existing equipment through low excess air operation or staged combustion discussed previously.

Under pressure of the New Source Performance Standards, boiler manufacturers and utility groups have developed burner designs that represent their approach to low NO_x emissions. This section will briefly discuss these designs and the available emissions data. Cost information is also included where available. In addition, discussions of burner modifications applied by Arizona Public Service for pulverized coal firing and of vertically fired boilers are included as they represent alternative approaches that have demonstrated low NO_x operation.

B&W Dual Register Burner

The Babcock and Wilcox (B&W) Company has developed a combustion system design that has been successfully applied to new units as well as a retrofit to existing coal-fired utility boilers to achieve low NO_x emission levels. The system consists of: (1) a controlled mixing burner; (2) compartmented windbox system; and (3) increased furnace size over pre-NSPS B&W units. All combustion air is introduced into the furnace at the burner to prevent corrosion resulting from reducing atmospheres in the furnace.

The key to the system is a Dual Register Burner that inhibits the formation of NO_x through controlled mixing of the fuel and air streams. A fuel-rich internal jet is formed with delayed mixing of the secondary combustion air (2-3). The coal combustion thereby approaches a diffusion type flame which has been shown to have lower NO_x emissions than the turbulent, nearly premixed flame characteristics of more conventional coal burner designs. The dual register design is reported by B&W to achieve low NO_x levels through reduced peak flame temperatures and controlled oxygen availability (2-20).

A low overall excess air level is maintained by precise control of the air and fuel flow to each burner. Coal flow through each burner pipe is monitored and adjusted to provide uniform distribution with burner matrix. Compartmented windboxes are included to balance combustion air flow.

The dual register burner was initially evaluated as a retrofit in a 270 MW new horizontally opposed unit. B&W Dual Register Burners have since been used on seven new units designed to meet NSPS.

Design. The major differences between the Dual Register and earlier B&W burners are the independent air registers for secondary and tertiary air flow control and adjustable swirl vanes in the secondary air lanes to control axial and tangential velocity, secondary air mixing patterns and ultimately the flame location. Optimization of the coal and air mixing patterns can therefore be made for low NO_x emissions.

The B&W Dual Register Burner was equipped with a venturi section in the coal nozzle to maintain a homogeneous particle distribution necessary for flame stability and complete carbon burnout. An adjustable plug (spear) within the venturi section provides control of the velocity and subsequent turbulence of the coal and primary air stream. Use of the venturi also results in a more axial flow pattern and thereby reduces mixing more than that achieved with the impeller. This design is shown in Figure 2-44.

Uniform primary air flow to each burner is maintained using calibrated orifices located in the coal pipes between the pulverizers and burners. Fine control is made using the movable plug in the venturi section shown in Figure 2-44.

Compartmented windboxes are used on new units to provide a uniform air supply to each burner's secondary and tertiary air registers. Air foils are also included to maintain desired air flow conditions. Overall minimum excess air levels are therefore achievable with associated low NO_x levels without excessive carbon carryover or CO emissions. B&W claims that the burner and windbox configuration of the dual register burner provides the user with the best overall air-fuel ratio control available for large-scale pulverized-coal systems.

Furnace designs have also been modified by increasing burner zone dimensions (2-37). Burner zone volumes for new units have been increased by 130 percent over 1960 designs. Associated increases in burner zone surface area have resulted in decrease in heat release rates per unit area by 40 percent. Additional flame cooling is achieved by increasing the spacing between burners and thereby reducing combustion intensity.

Field Test Results. As mentioned previously, the first field tests of the Dual Register Burner were conducted on a 270 MW opposed fired boiler, burning bituminous coal (2-37). The use of low excess air allowed the flame to approach the designed mixing patterns and achieved marked NO_x emission reductions. When properly adjusted, the new burners reduced the NO_x from 725 to 365 ppm (3 percent O_2), a 50 percent reduction from the unit's original burners. CO emissions were under 50 ppm for all tests and carbon losses were not found to be excessive. No other furnace-related problems were found during these tests.

NO emission levels were sensitive to both secondary and tertiary air register position as well as the secondary swirl vane position. For reduced secondary air

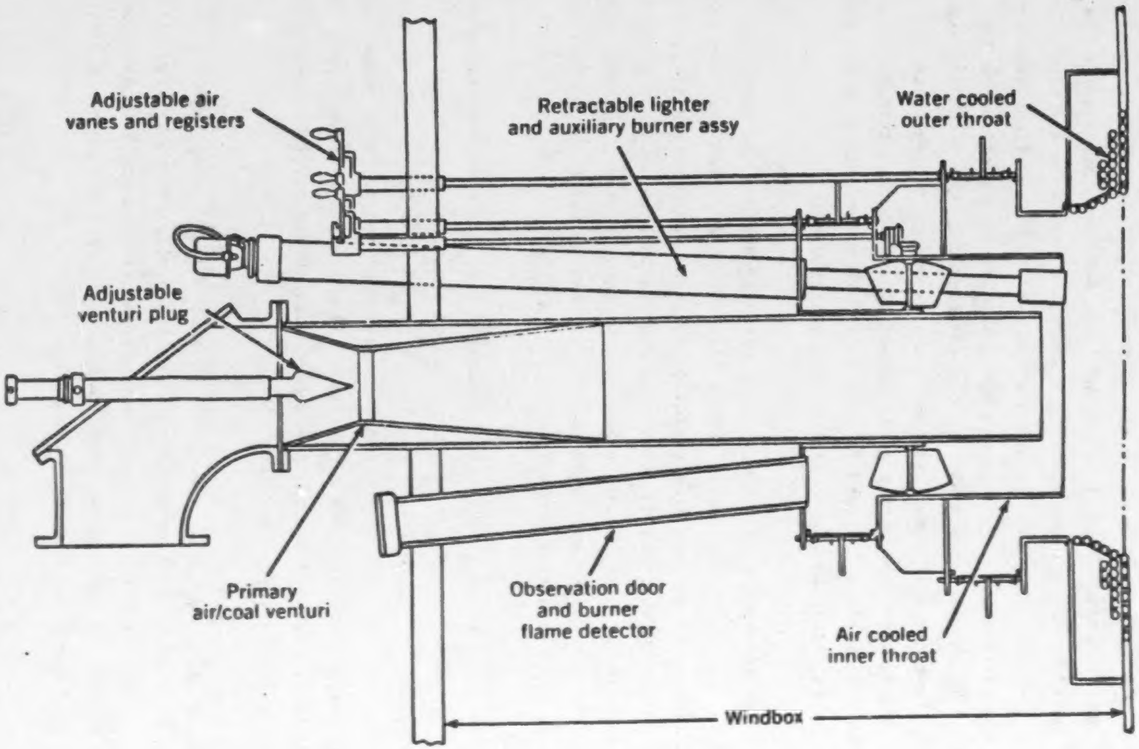
Adjustable air
vanes and registers
Retractable lighter
and auxiliary burner assy
Water cooled
outer throat
Adjustable
venturi plug
Primary
air/coal venturi
Observation door
and burner
flame detector
Air cooled
inner throat
Windbox

Figure 2-44. B&W Dual Register Burner Used On New Units (2-20).

flow, i.e., full open tertiary air register position, NO emissions decreased to a minimum at approximately 70 percent open position and then increased as the damper was moved to 100 percent open. For increased secondary air flow, i.e., with the tertiary air registers 50 percent open, an opposite effect occurred. NO emissions increased with increasing secondary air register position to the 50 percent to 70 percent open position, then decreased as the secondary air register was moved to full open. Adjusting the secondary air vanes to induce increasing counterclockwise spin also increased NO emissions. Evidently, the NO formation pattern in these burners is very sensitive to relatively small changes in combustion air distribution in the primary burner region.

Tests of the effectiveness of combined Dual Register Burners and burners-out-of-service (BOOS) operation were made. Using the conventional B&W circular burner, a NO_x reduction of 43 percent from as-found emissions was achieved through two-staged combustion. Resulting NO_x levels were comparable to those achieved without burners-out-of-service using the B&W Dual Register Burners. An additional reduction of 16 percent to approximately 400 ppm was achieved using combined dual-register and two-staged combustion operation (2-38).

Since the initial test series, B&W has installed Dual Register Burners on seven new units. The full load NO_x emissions data for these units at 15 percent excess air are presented in Figure 2-45. NO_x levels ranged from 250 to 275 ppm for subbituminous coals and 350 to 450 ppm for bituminous coals. Also presented in Figure 2-45 are the uncontrolled NO_x emissions for similar units equipped with circular burners. These data indicate a potential 40 percent to 60 percent reduction over previous B&W designs.

Application to Oil Combustion. The dual register burner concept has been modified to be used on utility-sized boilers firing oil. Shown in Figure 2-46 is the configuration for the oil burner. Significant differences from the coal burner are evident in the secondary and tertiary air flow patterns. A diffuser is used to hold the flame near the burner tip.

This burner has been used as a retrofit to utility-sized boilers to achieve NO_x emission levels of 225 ppm. With the conventional B&W combustion system used on these units, BOOS and flue gas recirculation were needed to achieve these levels.

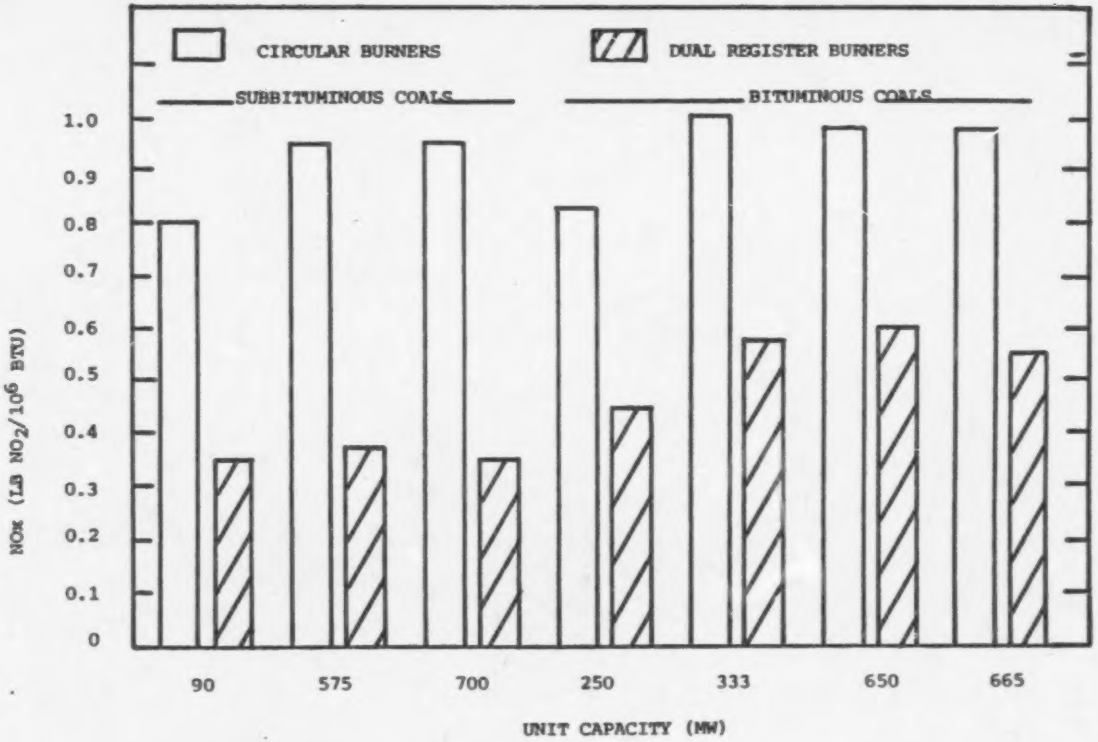

Figure 2-45. Comparison of NOx emissions for circular vs. B&W dual register burners at 15 percent excess air (2-37, 2-38).

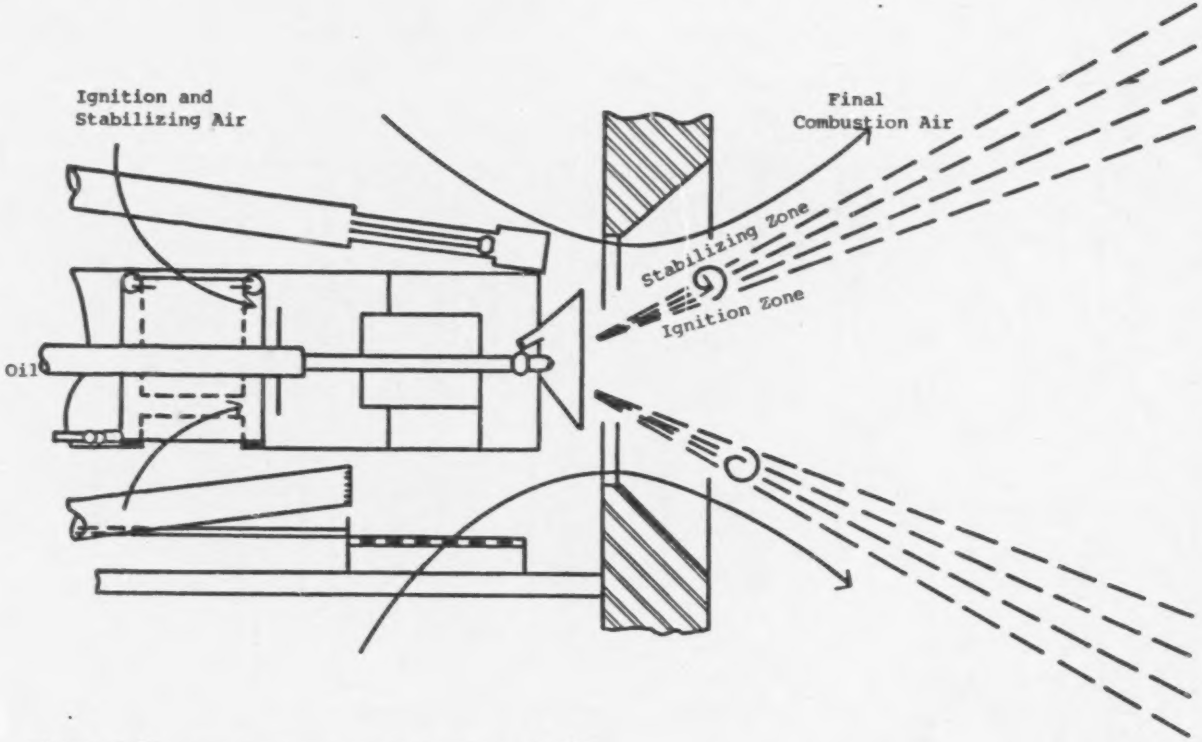
Ignition and
Stabilizing Air
Final
Combustion Air
Stabilizing Zone
Ignition Zone
Oil

Figure 2-46. B&W dual register oil burner (2-38).

Dual Register Burner Costs. Based on verbal discussions with B&W, the incremental investment costs of the dual register combustion system have been estimated to be ~$50 per kW (2-29). This cost includes the installation of the compartmented windbox and dual register burner. An annual cost of 0.03 mills per kW has been estimated for operation and maintenance of these systems (2-23).

Foster Wheeler Dual Register Burner

The Foster Wheeler Energy Corporation (FW) is in the process of developing and evaluating an advanced combustion system to achieve low NO_x emissions from pulverized coal combustion. A dual register burner has been developed using a configuration significantly different from the B&W design. Low NO_x emission levels have been achieved as compared with the conventional FW Intervane Burner when combined with overfire air operation (2-39). An advanced split flame burner has also been designed incorporating the primary features of the FW Dual register Burner with additional near burner staging.

Like the B&W Dual Register Burner, the FW combustion systems inhibit the formation of NO_x by controlling the mixing of the coal and combustion air streams. A fuel rich central jet is formed through controlled, delayed mixing of the secondary air stream. Means to control both the velocity of the coal/primary air and secondary air streams have been provided to allow optimal velocity difference (and thereby mixing) to be achieved. In the more advanced burner, the coal stream is further split into several coal jets which each burn somewhat independently within the main flame envelope. These burners reduce NO_x by having lower peak flame temperatures and controlled oxygen availability (2-39).

The FW systems provide balanced combustion air flow to each burner through burner hardware rather than compartmented windboxes as in the B&W systems. FW has also incorporated overfire air to further stage the combustion process. Curtain air injected along the furnace tube surface has been used to help alleviate slag formation and corrosion.

Design. Figure 2-47 shows the general configuration of the FW Dual Register Burner. The pulverized coal and primary air are supplied through two concentric pipes. An axially movable sleeve has been incorporated into the inner nozzle to control the velocity of the coal/primary air streams. In the split-flame burner

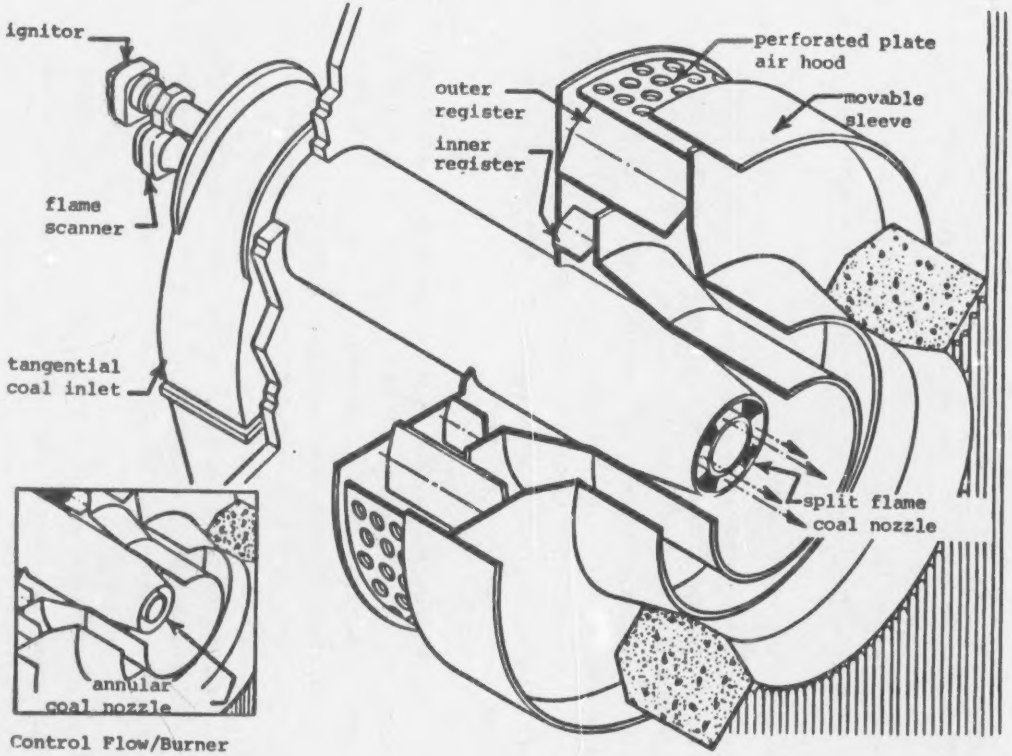
ignitor
flame scanner
tangential coal inlet
outer register
inner register
perforated plate air hood
movable sleeve
split flame coal nozzle
annular coal nozzle
Control Flow/Burner

Figure 2-47. Foster Wheeler low NO_x dual-register burner (2-40).

the coal flow is separated into four concentrated streams which form four separate flames. The mixing rate between the coal and primary air streams is thereby delayed resulting in rear burner flames of very low stoichiometry.

Secondary air flows through a perforated plate and an air register assembly before being divided into two secondary air lanes. Total flow through the perforated plate is controlled by a movable air hood which allows a balanced secondary air distribution between the burners to be achieved. The perforated plate assists in improving circumferential air distribution at each burner. An additional air register is located in the inner secondary air lane to control the distribution of secondary air at the burner face.

These burners have been evaluated alone and in combination with overfire air. Overfire air operation has been somewhat limited due to increased slag buildup in the lower furnace and around the burner throats.

FW has developed a curtain air concept in order to reduce this slagging and associated corrosion. Air is injected through slots located in several locations in the lower furnace and, as it passes up along the tube wall, coats the tube surface with an excess oxygen atmosphere. A reducing atmosphere can thereby exist in the central combustion region for reduced NO_x formation with an oxidizing atmosphere near the furnace walls for its protection.

Field Test Results. The FW Dual Register Burner has been evaluated as a retrofit on three units originally equipped with conventional FW Intervane Burners. The two larger units, rated at 265 MW, were also retrofitted with overfire air ports and curtain air. Both of these units are horizontally opposed with 12 burners on each wall, arranged in a 4 wide by 2 high matrix.

As shown in Figure 2-48, both overfire and the FW Dual Register Burners were shown to significantly reduce NO_x emissions. Operation with the original Intervane Burners and 100 percent overfire air reduced NO_x levels by 30 percent as compared to closed overfire air port operation. Prolonged continuous use of this procedure was prevented by increased slag buildup caused by reducing conditions in the furnace. It must be noted that the furnace was of Pre-NSPS design and was not, therefore, optimized for staged combustion.

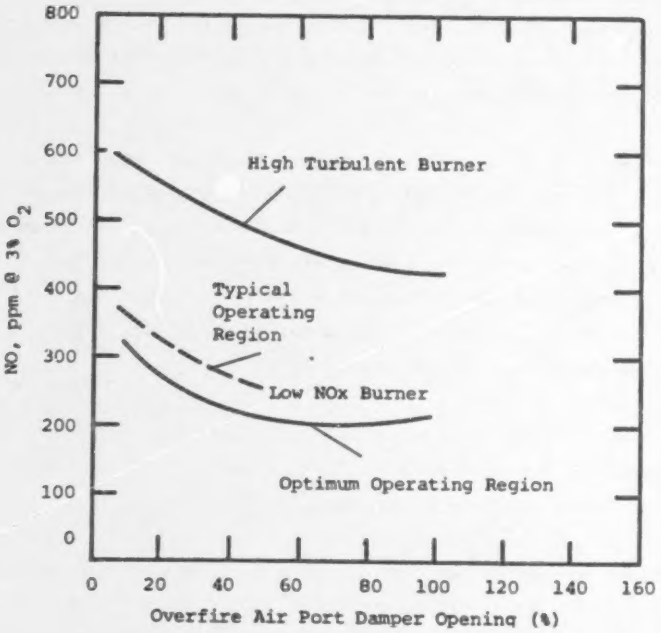

Figure 2-48. Comparison of NO_x emissions from Foster Wheeler Intervane and Dual Register Burners on 265 MW utility boiler (2-39).

NO_x emissions were reduced by nearly 50 percent using the FW Dual Register Burner to emission levels of 300 ppm. Emissions were further reduced to approximately 200 ppm by adding overfire air to this system. Total NO_x emissions were therefore reduced by 67 percent.

The curtain air added to these units with the Dual Register Burners substantially alleviated the slagging situation. Some slagging did persist due to the general furnace configuration.

The advanced FW Dual Register/Split Flame burner has been evaluated on a 125,000 lb/hr industrial sized boiler. NO_x emissions were reduced by 60 percent from comparable operations with FW Intervane Burners to a range of 200 to 250 ppm. The Dual Register Split Flame burner has been retrofitted on a 375 MW utility sized boiler. NO_x emission levels were below 0.4 lb/million Btu, representing a 65 percent reduction without the use of overfire air (2-40).

This Environmental Protection Agency (EPA) report evaluates and compares the operation of two EPA pilot plants, both reducing NOx emissions by 80-90% using Selective Catalytic Reduction (SCR). One process reduces NOx while the second process reduces both SOx and NOx. The simultaneous process is currently not as cost effective as using SCR and FGD together to control both SO2 and NOx.

Evaluation of NO_x and NO_x/SO_x Flue Gas Treatment Technology for Coal-Fired Sources

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Presented at

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Dallas, Texas
November 1-4, 1982

EPA'S PILOT PLANT EVALUATIONS OF NO_x AND NO_x/SO_x FLUE GAS TREATMENT TECHNOLOGY, J. David Mobley, U.S. Environmental Protection Agency, Industrial Environmental Research Laboratory, Research Triangle Park, NC 27711, and J. M. Burke, Radian Corporation, Austin, TX 78759.

Nitrogen oxide (NO_x) emissions from stationary sources may be reduced by 80 to 90 percent through the application of selective catalytic-reduction (SCR) of NO_x with ammonia (NH_3). For the most part, SCR has been developed and commercially applied in Japan. In the interest of evaluating SCR technology in the U.S., EPA has sponsored pilot-scale tests of two SCR systems treating flue gas from coal-fired boilers. The processes tested were the Hitachi Zosen process and the Shell Flue Gas Treatment (FGT) process, which also removes sulfur dioxide (SO_2). The results of those tests show that both processes tested are technically capable of reducing 90 percent of the NO_x in flue gas which contains the full particulate concentration from a coal-fired boiler; however, the performance of the Shell process was generally below this level. A sampling program conducted during the pilot plant tests showed no adverse secondary emissions from application of the SCR processes; however, high NH_3 and sulfur trioxide (SO_3) emissions are a concern due to their potential for causing operational problems in downstream equipment. Cost projections, which are based on pilot plant results, show that the Hitachi Zosen process combined with a conventional flue gas desulfurization process costs less than the Shell process for control of both NO_x and SO_2 emissions.

EPA'S PILOT PLANT EVALUATIONS OF NO_x AND NO_x/SO_x FLUE GAS TREATMENT TECHNOLOGY

INTRODUCTION

In order to assess pollution control technology, EPA has sponsored research in many areas including flue gas desulfurization, particulate control, and combustion modifications. This research takes many forms ranging from theoretical studies to large-scale tests of new pollution control processes and techniques.

EPA's research activities have also addressed selective catalytic reduction (SCR) of nitrogen oxides (NO_x) with ammonia. As part of an effort to assess this technology in the United States, the EPA has sponsored pilot plant tests of two SCR processes. The processes tested were the Hitachi Zosen process and the Shell Flue Gas Treatment (FGT) process, which also removes SO_2. The tests were conducted using flue gas from two coal-fired boilers, and both pilot plants processed about 1600 Nm^3/hr of flue gas which is equivalent to 0.5 MW of electric generating capacity.

The EPA-sponsored evaluations were the first tests of the Shell FGT process for NO_x and SO_2 removal on a coal-fired boiler and the first coal-fired tests of the Hitachi Zosen process in the United States. Although the Hitachi Zosen process has been tested on coal in Japan, EPA felt that differences between the composition and properties of coals burned in Japan and those burned in the United States could affect the performance of the catalyst. As the results showed, fly ash properties at the United States test sites did affect operation of the Hitachi Zosen process, but did not impact operation of the Shell FGT process.

OBJECTIVES

The SCR pilot plant tests had several objectives. The primary one was to document the ability of each process to treat flue gas from a coal-fired boiler. Other objectives included evaluation of the processes' ability to reduce NO_x emissions by 90 percent for 90 days; measurement of emissions from the catalytic reactor of pollutants such as NH_3, SO_3, hydrocarbons, and CO; and measurement of the raw material and utility requirements of each process. Test programs were developed to meet these objectives, and the pilot plants were operated for nearly 2 years. During this period, several catalysts were tested over a wide range of operating conditions.

PROCESS DESCRIPTIONS

Hitachi Zosen Process

The Hitachi Zosen process uses ammonia (NH_3), added to the flue gas, to selectively reduce NO_x to molecular nitrogen (N_2) in a catalytic reactor. The overall chemical reactions are shown in Equations 1 and 2.

$$4\ NO(g) + 4\ NH_3(g) + O_2(g) \rightarrow 4\ N_2(g) + 6\ H_2O(g) \qquad (1)$$

$$2\ NO_2(g) + 4\ NH_3(g) + O_2(g) \rightarrow 3\ N_2(g) + 6\ H_2O(g) \qquad (2)$$

In the process, NH_3 is injected into flue gas which is at a temperature between 300°C and 400°C. The gas then passes through a catalytic reactor where the reactions (Equations 1 and 2) occur rapidly.

The pilot test unit of the Hitachi Zosen process was located at Georgia Power Company's Plant Mitchell Station in Albany, Georgia. The Hitachi Zosen pilot plant processed a 1600-2400 Nm^3/hr (1000-1500 scfm) slipstream of flue gas which was withdrawn between the economizer and the air preheater of the coal-fired Unit 3 at the Plant Mitchell Station. At this point, the flue gas contained the full particulate loading from the boiler. The slipstream from the boiler was first passed up through a flue gas heater for temperature control. NH_3 was then injected into the flue gas which flowed down through the reactor. After exiting the reactor, the gas passed through a cyclone particulate collector, a fan, and then returned to the boiler flue gas duct. The cyclone was included in the pilot unit to minimize abrasion in the fan.

Shell Flue Gas Treatment Process

The Shell FGT process differs in that it removes both SO_2 and NO_x. Copper supported on alumina is used as an SO_2 "acceptor." SO_2 and O_2 react with the acceptor as shown in Equation 3.

$$SO_2(g) + 1/2\ O_2(g) + CuO(s) \rightarrow CuSO_4(s) \qquad (3)$$

Copper oxide and, to a greater extent, the copper sulfate ($CuSO_4$) produced by the SO_2 acceptance reaction act as a catalyst for the reduction of NO_x by the injected NH_3 according to Equations 1 and 2.

As SO_2 is removed in the reactor, the acceptor becomes "loaded," and SO_2 emissions begin to increase. Once the time averaged SO_2 removal reaches the design level (i.e., 90 percent), flue gas flow to the reactor is shut off and the $CuSO_4$ and CuO are reconverted to Cu metal. In this regeneration step, a steam/hydrogen mixture is passed through the reactor, and a concentrated SO_2 stream is produced. The regeneration reactions are illustrated by Equations 4 and 5.

$$CuSO_4(s) + H_2(g) \rightarrow CuO(s) + SO_2(g) + H_2O(g) \qquad (4)$$

$$CuO(s) + H_2(g) \rightarrow Cu(s) + H_2O(g) \qquad (5)$$

The SO_2 produced by the regeneration of the acceptor can be further processed to recover the sulfur as sulfuric acid or elemental sulfur.

The Shell FGT pilot test unit was located at Tampa Electric Company's Big Bend Station in North Ruskin, Florida. The pilot plant processed a 1000 scfm (1600 Nm^3/hr) slipstream of flue gas which was withdrawn between the economizer and the air preheater of the coal-fired Unit 2. Flue gas from the boiler contained full particulate loading and was heated by in-line burners before entering the reactor. Flue gas flowed up through the reactor during acceptance of SO_2 and bypassed the reactor during the regeneration of the acceptor. During the program, some tests were conducted in which the reactor was fully loaded with SO_2, and NO_x was the only pollutant removed. While these tests were conducted, the reactor processed flue gas continuously.

RESULTS(1,2)

The results of the pilot plant tests and subsequent data analysis can conveniently be divided into several categories, including: process performance, process energy requirements, process costs, and the flue gas impacts of the process. When added together, these results are indicative of the technical and economic feasibility considerations of applying these two SCR processes to coal-fired boilers. In the discussion which follows, the pilot plant test results are summarized separately for each of these categories.

Process Performance

The performance of the two pilot plants can best be summarized by the results of tests which were conducted to document the ability of the processes to reduce NO_x emissions by 90 percent over 90 days of continuous operation. In the case of the Shell FGT process, simultaneous SO_2 removal efficiency of 90 percent was also required.

Hitachi Zosen Process Performance—

The test results from the Hitachi Zosen pilot plant are presented in Figure 1. These results cover the period of May through October 1980, and represent over 3500 hours of operation. As illustrated by Figure 1, the Hitachi Zosen pilot plant met and exceeded the objectives of the test. During the first 90 days of operation, NO_x removal averaged over 90 percent. And, for the entire period of 160 days of continuous operation, the average NO_x reduction was 89.8 percent which is only slightly below the objective of 90 percent. Figure 1 also shows that in the range of 2100 to 2400 Nm^3/hr flow rate has essentially no effect on the NO_x removal efficiency of the process.

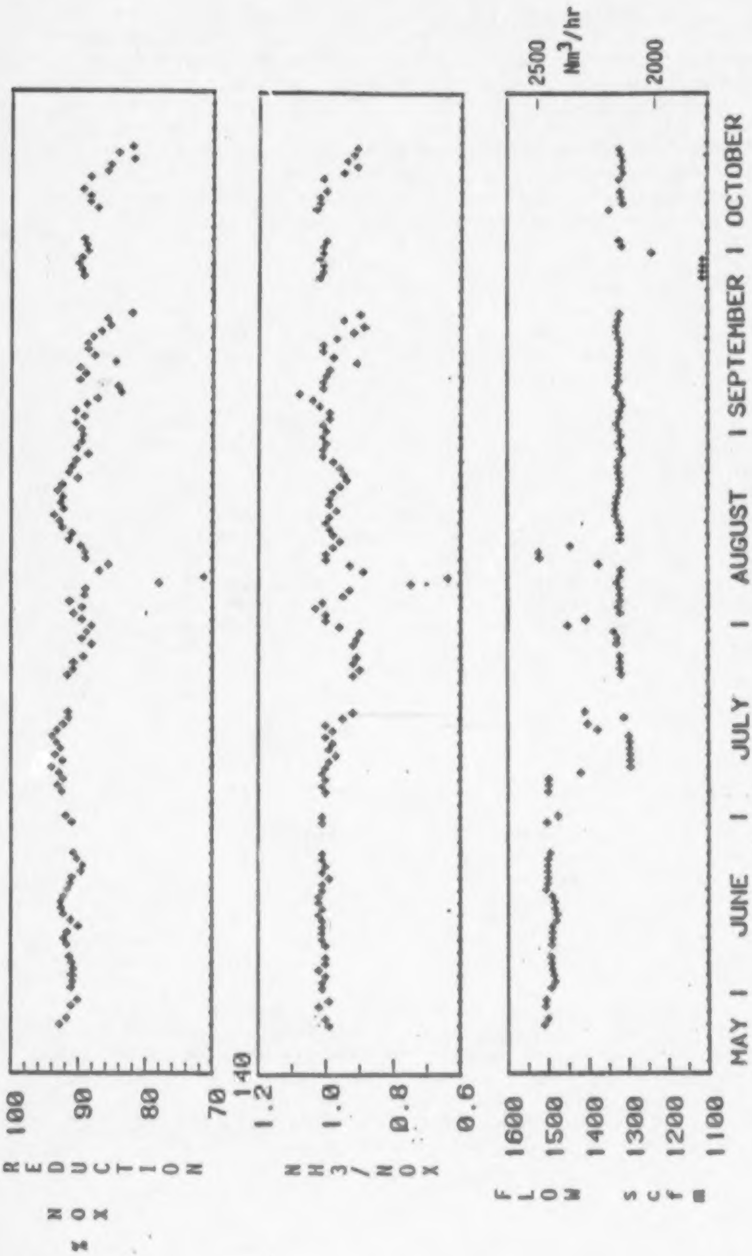

Figure 1. Pilot plant test results from the Hitachi Zosen process using NH_3 addition for NO_x reduction (1980).

The results presented in Figure 1 indicate several things. They show that the Hitachi Zosen process is capable of treating flue gas from a coal-fired boiler for long periods of time while maintaining high NO_x reduction efficiencies. But they also show that the activity of the catalyst gradually declined over the period May through October. This indicates that additional catalyst may be required to maintain 90 percent reduction over a 1 year period of operation. However, a catalyst regeneration technique was tested which may make it possible to maintain 90 percent NO_x reduction without additional catalyst. Near the end of the test program, the catalyst was regenerated using an in situ, hot water washing technique. Following regeneration, the catalyst performance was essentially identical to that of fresh catalyst. Unfortunately, it was not possible to determine how long the effects of regeneration last or how many times the catalyst can be regenerated before a permanent loss of catalytic activity occurs. As a result, it is still uncertain if catalyst regeneration will permit the process to operate for 1 year at 90 percent NO_x reduction efficiency without the use of additional catalyst.

The results presented in Figure 1 are for Zosen's NOXNON 600 series catalyst. Prior to testing the 600 series catalyst, the pilot plant tested the 500 series catalyst which has smaller openings for gas passage. During the tests, the NOXNON 500 catalyst became plugged with fly ash after about 2000 hours of operation. This was evidenced by an increase in reactor pressure drop and a decline in NO_x reduction efficiency, and it was believed to result from the adhesiveness of the fly ash at high temperatures. Tests were conducted which compared fly ash properties from a power plant in Japan with the properties of fly ash from the Plant Mitchell power plant in the United States. At reactor operating temperatures, fly ash from Japan remained as a fine powder, but fly ash from Plant Mitchell agglomerated to form large lumps of material. This indicates a significant difference between a Japanese and United States coal-fired power plant and illustrates the need for test work in the United States.

In summary, the Hitachi Zosen tests indicated the need for a catalyst with larger gas passages to prevent catalyst plugging when treating flue gas from a coal-fired power plant in the United States. The tests did demonstrate that the process is capable of maintaining high NO_x reduction efficiencies for long periods of time. And, although catalyst activity gradually declined during the test program, it may be possible to maintain 90 percent NO_x reduction efficiency for a 1 year period with appropriate process design and operating considerations.

Shell FGT Process Performance—

The test results from the Shell FGT pilot plant are shown in Figure 2. These results are for tests conducted during June 1980, and they represent approximately 300 hours of operation in which flue gas was processed in the reactor. As Figure 2 illustrates, the Shell FGT process did not meet the objectives of the demonstration test. Although the SO_2 removal efficiency averaged 90 percent during the test, the NO_x reduction averaged only 70 percent. And the test lasted only 20 days rather than the scheduled 90 days.

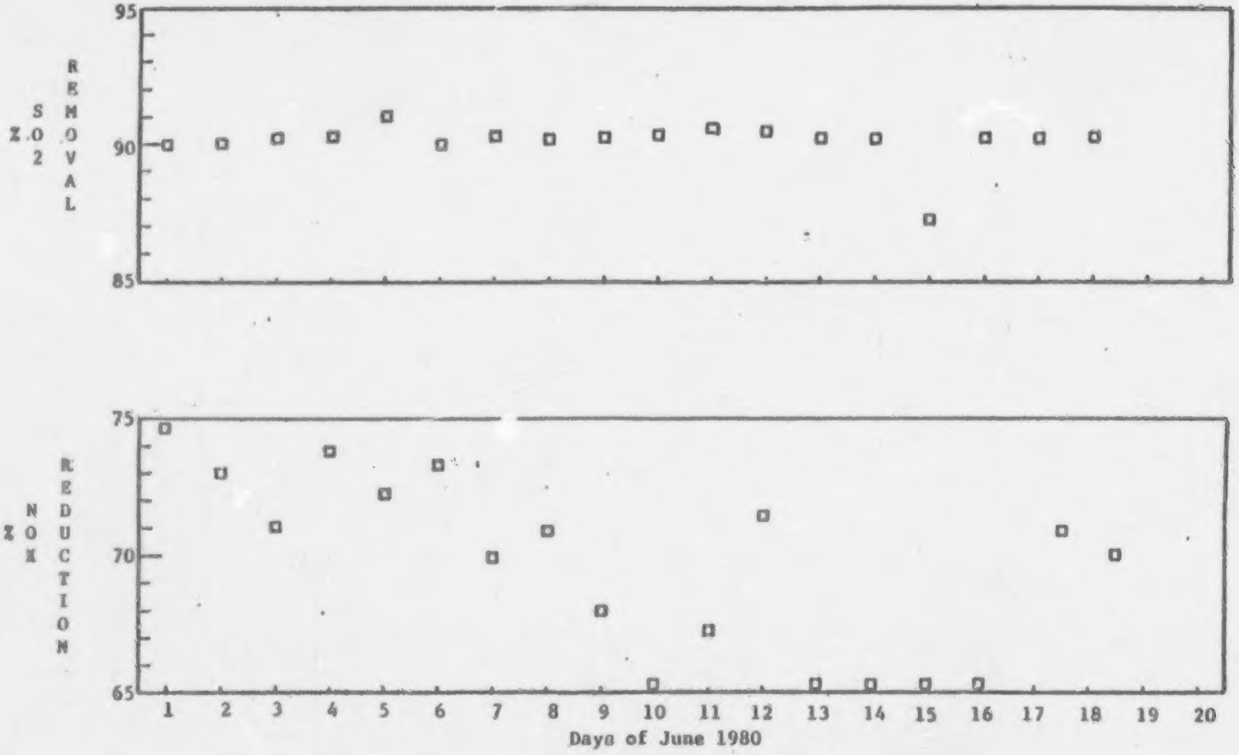

Figure 2. Pilot plant test results from the Shell Flue Gas Treatment process for SO_2 / NO_x removal from flue gas.

The inability of the pilot plant to meet the test objective of simultaneously reducing NO_x and SO_2 emissions by 90 percent was due to the poor performance of the catalyst used in the process. With the reactor configuration used during the pilot plant tests, the NO_x reduction efficiency was low even when NH_3 was injected to give an NH_3 to NO_x ratio of 1.3. The NO_x reduction efficiencies of the pilot plant could be improved, but only at the expense of reduced SO_2 removal.

The results of the tests at the Shell FGT pilot unit did not demonstrate simultaneous reduction of NO_x and SO_2 emissions by 90 percent. However, the pilot plant test results can be extrapolated to identify conditions which will result in 90 percent reduction of both NO_x and SO_2 emissions. To achieve this level of emissions reduction, additional catalyst would have to be added to the reactor. Under the conditions at the pilot plant test site, 50 to 75 percent additional catalyst would have been required. While addition of catalyst is technically feasible, it would significantly increase the costs of applying the process. (An alternative method of improving NO_x reduction efficiency of the Shell FGT process was identified during the test program. The alternative method involved addition of processing steps to the SO_2-removal/NO_x-reduction cycle of the reactor. In a commercial design, the additional processing steps would require additional process equipment. This evaluation of the Shell FGT process did not consider the alternative method since sufficient information was not available to enable a cost estimate; however, the reader should be aware that the alternative method may be less costly than use of additional acceptor in the reactor.)

Another thing which was not attained during the tests with the Shell FGT process was the stability of the catalyst activity over a long period of time. Three catalysts were tested; of the three, the second processed flue gas for the longest period of time. Examination of the activity of this catalyst, as evidenced by its SO_2 removal efficiency, indicated that activity dropped after the catalyst's initial exposure to flue gas. The catalyst activity then stabilized and no measurable change in activity occurred over a 700 hour period. While this indicates that the catalyst activity appears stable for a coal-fired application, flue gas was not processed long enough to reach any conclusions.

Perhaps the most significant result of the tests with the Shell process was that no plugging of the reactor occurred. Flue gas was processed for a total of about 2000 hours and no increase in reactor pressure drop was recorded. This indicates that the parallel passage reactor design used in the pilot plant tests is well suited for application to coal-fired sources.

In summary, the tests of the Shell FGT process did not meet the objectives of the process to reduce emissions of both NO_x and SO_2 by 90 percent. However, the results can be extrapolated to identify a process design which will reduce NO_x and SO_2 emissions 90 percent. The test results did indicate that the parallel passage reactor design is well suited to coal-fired applications since no evidence of reactor plugging was found. But the test did not assess the long-term stability of the catalyst activity.

Process Energy Requirements

One of the key factors used by the EPA in evaluating any pollution control process is the fraction of the boiler's output which is required to operate the pollution control system. This is a measure of how much of the boiler's capacity does not generate electricity for sale and is thus one factor which contributes to the costs of applying a pollution control process.

As part of an evaluation of the pilot plant test results, the energy requirements of both the Hitachi Zosen and the Shell FGT processes were estimated. These estimates were made for a 500 MW application of each process, and they were based on measurements made during the tests as well as estimates made in a previous study.(3)

Since the Shell FGT process reduces emissions of both NO_x and SO_2 while the Hitachi Zosen process reduces only NO_x emissions, a direct comparison of process energy requirements is not meaningful. Instead, energy requirements for the Hitachi Zosen process and energy requirements for a flue gas desulfurization process were estimated and compared with the energy requirements for the Shell FGT process. The results of these estimates are presented in Table I, which identifies individual components and the overall process energy requirements as a percentage of the boiler's capacity.

As Table I shows, the Shell FGT process has an energy requirement equivalent to 5 percent of the boiler's capacity while the Hitachi Zosen combined with a flue gas desulfurization system required only 3.7 percent of the boiler's capacity. The principal difference between the energy demands of the two processes is the fuel energy requirement of the Shell FGT process.

TABLE I. COMPARISON OF HITACHI ZOSEN AND SHELL FLUE GAS TREATMENT PROCESS ENERGY REQUIREMENTS[a,b]

Energy Component	Hitachi Zosen[c]	Shell
Fuel	-	5.6
Steam	2.0	0.6
Electricity	2.0	2.0
Energy Credit	(0.3)	(3.2)
Total	3.7	5.0

[a] All energy requirements as percent of boiler capacity.

[b] 500 MW coal-fired boiler. Boiler thermal efficiency = 88%. Boiler heat rate = 9.5 MJ/kWh.

[c] Includes energy requirements for limestone flue gas desulfurization system.

The fuel energy consumed by the Shell process is equivalent to 5.6 percent of the boiler's capacity, and it is actually greater than the total process energy requirements. To some extent, the high fuel energy requirement of the Shell process is offset by the energy credit which results from application of the process. The energy credit is primarily the result of additional heat recovery in the air preheater due to lower permissible operating temperatures. Additional heat recovery is possible since samples collected during the pilot plant tests indicated that sulfur trioxide is removed from the flue gas in the reactor. This lowers the temperature at which sulfuric acid condenses (dew point), thus reducing the chance of corrosion in, or downstream of, the air preheater.

Another item of interest which is not shown in Table I is that the majority of the energy requirements shown for the Hitachi Zosen process are those of the flue gas desulfurization system: they are not associated directly with the SCR process. The estimated energy requirements of the Hitachi Zosen process alone are equivalent to only 0.3 percent of the boiler's capacity. These energy requirements result from compressed air and steam used for soot blowing and the electrical energy required to overcome the reactor pressure drop.

Process Costs

Estimates were prepared of total capital investment and annual revenue requirements for the Hitachi Zosen process and the Shell FGT process based on results from the pilot plant tests and estimates from a previous study.(3) These costs reflect the way each process was operated during the pilot plant test program, and they can be used to compare the two processes. However, since the Hitachi Zosen process does not remove SO_2, the costs of a flue gas desulfurization system(3) must be added to the Hitachi Zosen process cost estimate to make comparison with the Shell FGT process costs meaningful.

Table II contains the results of the cost estimates for the Hitachi Zosen and Shell processes and the cost estimate for a limestone flue gas desulfurization system. As shown, the Shell process costs are estimated to be significantly higher than the costs for the Hitachi Zosen process, including the costs of flue gas desulfurization. The capital investment for the Shell process is estimated to be $84.2 million or $168/kW of generating capacity. The capital investment for the Hitachi Zosen process including flue gas desulfurization is estimated to be $72.5 million or $145/kW of generating capacity. This indicates that the capital investment of the Shell process is over 15 percent higher than that for the Hitachi Zosen process. The primary reason for the Shell FGT process's being more expensive is the relatively poor NO_x reduction efficiencies attained by the process during the pilot plant tests.

TABLE II. ESTIMATED CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS FOR A 500 MW APPLICATION OF THE HITACHI ZOSEN AND SHELL FLUE GAS TREATMENT PROCESSES[a,b]

Cost Item	Hitachi Zosen Process			Shell Process
	HZ	FGD	Total	
Capital Cost	\$22,100,000	\$50,400,000	\$72,500,000	\$84,200,000
Annual Revenue Requirement	10,200,000	14,700,000	24,900,000	33,500,000

[a] A detailed breakdown of the Hitachi Zosen and Shell process cost estimates is included as an appendix to this paper.

[b] Basis for Estimates: 500 MW new coal-fired power plant burning 3.5 percent sulfur coal. Removal of both NO_x and SO_2 = 90 percent. Plant location in Midwestern United States. Project beginning mid-1977, ending mid-1980. Remaining life of power plant = 30 years. Plant on line 7000 hr/yr. Plant heat rate = 9.5 MJ/kWh.

Perhaps a more important measure of relative process costs is the annual revenue requirements of the processes. The revenue requirements represent the annual costs of process operation including raw material and energy requirements, depreciation, return on investment, and catalyst replacement. These revenue requirements are also presented in Table II. It should be noted that these costs are based on the assumption of a 1 year life for the catalysts used in both processes. As discussed earlier, this assumption cannot be verified by the pilot plant test results, but a 1 year catalyst life is typicallly guaranteed by process vendors.

As shown in Table II, the annual revenue requirements for the Shell process are estimated to be \$33.5 million/year. This is over 30 percent greater than the annual revenue requirements for the Hitachi Zosen process which were estimated to be \$24.9 million/year. Again, the principal reason for the difference between the annual revenue requirements for the two processes is the relatively poor performance exhibited by the Shell process during the pilot plant tests which indicated that a substantial increase in catalyst volume would be required to achieve 90 percent NO_x removal. Another significant difference is the relatively high energy requirements of the Shell process.

Overall, the results of the cost estimates indicate that the Shell FGT process requires a greater capital investment and has higher annual revenue requirements than the Hitachi Zosen process. However, the cost estimate was prepared only for a 500 MW coal-fired application with the requirements of 90 percent SO_2 and NO_x removal. The relative costs for the two processes could differ for other applications with different requirements.

Process Flue Gas Impacts

As part of the pilot plant tests, EPA conducted a sampling program designed to quantify process emissions of pollutants other than NO_x and SO_2. The objective of this sampling program was to determine if any adverse flue gas concentration impacts can result from application of SCR technology to a coal-fired source. Samples were collected at the inlet and outlet of the pilot plant reactors and analyzed for selected pollutants. Inlet and outlet samples were collected simultaneously, so that measured differences in pollutant concentration were due to changes across the reactors.

Table III identifies the specific pollutants measured and presents the results of the sampling program. As shown, the concentrations of some pollutants changed across the reactor while others did not. And in some cases, the concentrations of the pollutants were below the detection limits of the analytical techniques employed during the sampling program.

TABLE III. MEASURED CHANGE IN THE CONCENTRATIONS OF SELECTED POLLUTANTS ACROSS THE PILOT PLANT SCR REACTORS

	Hitachi Zosen Pilot Plant		Shell Flue Gas Treatment Pilot Plant	
Pollutant	Reactor Inlet Concentration	Reactor Outlet Concentration	Reactor Inlet Concentration	Reactor Outlet Concentration
Hydrogen Cyanide[a]	<10 $\mu g/m^3$	<10 $\mu g/m^3$	<10 $\mu g/m^3$	<10 $\mu g/m^3$
Nitrosoamines[a]	<5 $\mu g/m^3$	<5 $\mu g/m^3$	<5 $\mu g/m^3$	<5 $\mu g/m^3$
Carbon Monoxide	<0.017 %	<0.017 %	0.13 %	<0.017 %
Hydrocarbons (C_1-C_6)	<1.0 ppmv	<1.0 ppmv	28.5 ppmv	21.0 ppmv
Sulfur Trioxide	8.4 ppmv	20.7 ppmv	11.4 ppmv	0.1 ppmv
Ammonia	0	54.8 ppmv	0	15.3 ppmv

[a] Gas volume taken at 15.5°C, 1 atm, dry basis.

As illustrated in Table III, concentrations of hydrogen cyanide and nitrosoamines at the inlet and outlet of the pilot plant reactors were below the detection limits of the analytical techniques employed. In both cases, these concentrations are below the level which is considered safe for emissions from sources.

At the Hitachi Zosen plant, the concentrations of both CO and hydrocarbons were also below the detection limits of the analytical techniques. But at the Shell FGT pilot plant, the concentrations of these compounds were found to decrease across the reactor. This decrease represents an environmental benefit for the Shell process, and is believed to result from oxidation of these compounds in the reactor. It should be noted that the Hitachi Zosen reactor may also oxidize CO and hydrocarbons, but no conclusions can be drawn since these compounds were present in such low concentrations.

One of the most significant results of the emissions sampling program was the effect of the Hitachi Zosen and Shell processes on the concentration of SO_3 in the flue gas. As shown in Table III, SO_3 was produced in the Hitachi Zosen pilot plant reactor while the Shell FGT process removed SO_3 from the gas stream. These results are significant because of the effects SO_3 can have on equipment located downstream of an SCR reactor, especially the air preheater.

In the case of the Hitachi Zosen process, the increased SO_3 concentration raises the sulfuric acid dew point of the flue gas. This increases the potential for formation of ammonium sulfates. The net effect of increased flue gas SO_3 concentrations is to increase process costs. This is reflected in the cost estimates of air preheater modifications which include provisions for soot blowing, water washing, and periodic replacement of the air preheater elements.

In the case of the Shell process, removing SO_3 reduces the sulfuric acid dew point of the flue gas. This permits the air preheater to operate at lower temperatures without experiencing corrosion in the preheater or in downstream equipment. The result is additional heat recovery which lowers process energy requirements and thus process costs. The low SO_3 concentration at the reactor outlet also minimizes the potential for formation of ammonium sulfates.

Another significant result of the emissions sampling program is the measured NH_3 emissions from the processes. As shown in Table III, NH_3 emissions from the Hitachi Zosen process were over three times greater than the emissions from the Shell FGT process even though the NH_3/NO_x injection ratio was higher at the Shell pilot plant. This is due to the fact that the catalyst used in the Shell process promotes NH_3 oxidation while essentially no NH_3 oxidation occurred in the Hitachi Zosen process.

These results are significant since they indicate the potential for formation of ammonium sulfates downstream of the SCR reactors. Formation of ammonium sulfate is significant because it can result in corrosion and plugging of the air preheater and thus increase boiler operating costs. In the case of the Hitachi Zosen process, high NH_3 and SO_3 emissions indicate that a large quantity of ammonium sulfates can form downstream of the reactor. The Shell FGT process, on the other hand, has low emission rates of both NH_3 and SO_3. Thus, the potential for formation of ammonium sulfates is low.

CONCLUSIONS

Several important conclusions can be drawn from the results of the pilot plant test program. First, and most important, the test results show that both the Hitachi Zosen NO_x removal process and the Shell NO_x/SO_x control process are technically capable of achieving significant reductions of emissions from coal-fired boilers in the United States. Hitachi Zosen's NOXNON 600 catalyst and the Shell parallel passage reactor operated without fly ash plugging problems during the test program. The tests did show, however, that Hitachi Zosen's 500 series catalyst is prone to plugging. In contrast, tests by Hitachi Zosen in Japan reported that the NOXNON 500 catalyst treated flue gas from a coal-fired boiler for over 5000 hours without plugging problems. This difference between the properties of Japanese and United States coals points out the need for SCR test work when considering SCR process applications for untested coals.

In terms of performance, the Hitachi Zosen process met the test program objectives of 90 percent NO_x reduction for 90 days while the Shell FGT process did not. The Shell process did not meet the test program objectives because the catalyst performance was below expected levels and, therefore, the pilot plant reactor was undersized.

The pilot plant tests did not demonstrate a 1 year catalyst life, but the Hitachi Zosen catalyst did process flue gas for 6 months while averaging nearly 90 percent NO_x reduction. And although the Hitachi Zosen 600 series catalyst activity did decline during the test program, a regeneration technique was developed which restored catalyst activity to its original level. The performance of the Shell catalyst was quite stable during the test program, but the catalyst only processed flue gas for an equivalent of 3 months of operation. In both cases, a catalyst life of 1 year is typically guaranteed by the process vendors.

A comparison of the two processes, based on certain assumptions, shows that the Hitachi Zosen process combined with a flue gas desulfurization system has lower energy requirements and lower costs than the Shell FGT process. On the negative side, however, the Hitachi Zosen process has higher NH_3 and SO_3 emissions. Even so, it appears that the Hitachi Zosen process costs less than the Shell FGT process for a 500 MW boiler burning high sulfur coal.

The pilot plant tests show that it may be possible to achieve high NO_x and NO_x/SO_x control by use of FGT technology on coal-fired sources in the United States. However, a prototype-scale test on a 10 to 100 MW facility may be necessary before the technology can be considered demonstrated for coal-fired sources in the United States.

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APPENDIX

CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS FOR FULL SCALE SYSTEMS

TABLE A-I. ESTIMATED CAPITAL INVESTMENT FOR A 500 MW APPLICATION OF THE SHELL FGT PROCESS[a]

Item	Investment, $	% of total direct investment
Direct Investment[b]		
NH_3 storage and injection	890,000	1.9
H_2SO_4 plant	7,172,000	15.2
Reactor section	27,455,000	58.1
Flow smoothing section	2,828,000	6.0
Steam naphtha reformer	4,698,000	9.9
Gas handling	1,546,000	3.3
Sub-total direct investment (DI)	44,589,000	94.4
Services, utilities (0.06) (DI)	2,675,000	5.6
Total direct investment (TDI)	47,264,000	100.0
Indirect Investment		
Engineering design and supervision	709,000	1.5
Architect and engineering contractor	177,000	0.4
Construction expense = 0.25 $(TDI \times 10^{-6})^{0.83}$	6,135,000	13.0
Contractor fees = 0.096 $(TDI \times 10^{-6})^{0.76}$	1,799,000	3.8
Total indirect investment (IDI)	8,820,000	18.7
Contingency = 0.2 (TDI + IDI)	11,217,000	23.7
Total fixed investment (TFI)	67,301,000	142.4
Other Capital Charges		
Allowance for start-up and modifications = (0.1) (TFI)	6,730,000	14.2
Interest during construction = (0.12) (TFI)	8,076,000	17.1
Total depreciable investment	82,107,000	173.7
Land	14,000	—
Working capital	2,051,000	4.4
TOTAL CAPITAL INVESTMENT	84,172,000	178.1

[a] Basis: 500 MW new coal-fired power plant, 3.5% sulfur coal, 90% NO_x removal, 90% SO_2 removal. Midwest plant location. Represents project beginning mid-1977, ending mid-1980. Average basis for scaling, mid-1979. Investment requirements for fly ash disposal excluded. Construction labor shortages with overtime pay incentive not considered.

[b] Each item of direct investment includes total equipment costs plus installation labor, and material costs for electrical, piping, ductwork, foundations, structural, instrumentation, and site preparation.

TABLE A-II. ESTIMATED AVERAGE ANNUAL REVENUE REQUIREMENTS FOR A 500 MW APPLICATION OF THE SHELL FGT PROCESS[a]

Item	Annual quantity	Unit cost, $	Annual cost, $	% of annual revenue required
Direct Costs				
Raw materials				
NH_3	7.875x10^6 kg	0.165/kg	1,299,500	3.88
Naphtha	39.773 m^3	132.1/m^3	5,254,000	15.69
Catalyst	—	—	12,600,300	37.63
Reformer catalyst	—	—	125,100	0.37
Total raw materials			19,278,900	57.57
Conversion costs				
Operating labor and supervision	29,200 labor hrs	12.50/labor hr	365,000	1.09
Utilities				
Naphtha	9,943 m^3	132.1/m^3	1,313,500	3.92
Steam	179,900 GJ	-1.896/GJ	341,200	1.02
Process water	8,078,000 m^3	0.01585/m^3	128,000	0.38
Electricity	68,859,000/kWh	0.029/kWh	1,996,900	5.96
Heat credit	1,047,500 GJ	-1.896/GJ	(1,986,000)	(5.93)
Maintenance			1,891,000	5.65
Analyses	4,380 labor hrs	17.00/labor hr	74,500	0.22
Total conversion costs			4,124,100	12.31
Total direct costs			23,403,000	69.88
Indirect Costs				
Capital charges				
Depreciation = (0.06) (total depreciable investment)			4,926,400	14.71
Average cost of capital = (0.086) (total capital investment)			7,238,800	21.62
Overheads				
Plant = (0.5) (conversion costs minus utilities)			1,165,300	3.48
Administrative = (0.1) (operating labor costs)			36,500	0.11
Marketing = (0.1) (sales revenue)			366,000	1.09
Total indirect costs			13,733,000	41.01
Spent catalyst disposal			11,400	0.03
Gross average annual revenue requirement			37,147,400	110.92
Byproduct Sales Revenue				
H_2SO_4	1.11x10^8 kg	-0.033/kg	(3,660,000)	(10.92)
TOTAL ANNUAL REVENUE REQUIREMENTS			33,487,400	100.00

[a] Basis: 500 MW new coal-fired power plant, 3.5% S coal. 90 percent NO_x reduction, 90 percent SO_2 removal. Midwest power plant location, 1980 revenue requirements. Remaining life of power plant = 30 years. Plant on line 7000 hr/yr. Plant heat rate = 9.5 GJ/kWh. Investment and revenue requirement for disposal of fly ash excluded. Total direct investment $47,264,000; total depreciable investment $82,107,000; and total capital investment $84,172,000.

TABLE A-III. ESTIMATED CAPITAL INVESTMENT FOR A 500 MW APPLICATION OF THE HITACHI ZOSEN PROCESS[a]

Item	Investment, $	% of total direct investment
Direct Investment[b]		
NH_3 storage and injection	645,000	5.5
Reactor section	8,632,000	73.4
Gas handling	351,000	3.0
Air preheater modifications	1,461,000	12.4
Sub-total direct investment (DI)	11,089,000	94.3
Services, utilities (0.06) (DI)	665,000	5.7
Total direct investment (TDI)	11,754,000	100.0
Indirect Investment		
Engineering design and supervision	274,000	2.3
Architect and engineering contractor	69,000	0.6
Construction expense = 0.25 $(TDI \times 10^{-6})^{0.83}$	1,933,000	16.4
Contractor fees = 0.096 $(TDI \times 10^{-6})^{0.76}$	625,000	5.3
Total indirect investment (IDI)	2,901,000	24.7
Contingency = 0.2 (TDI + IDI)	2,931,000	24.9
Total fixed investment (TFI)	17,586,000	149.6
Other Capital Charges		
Allowance for start-up and modifications = (0.1) (TFI)	1,759,000	15.0
Interest during construction = (0.12) (TFI)	2,110,000	17.9
Total depreciable investment	21,455,000	182.5
Land	5,000	—
Working capital	336,000	2.9
Royalty fee	300,000	2.6
TOTAL CAPITAL INVESTMENT	22,096,000	188.0

[a] Basis: 500 MW new coal-fired power plant, 3.5% sulfur coal, 90% NO_x removal. Midwest plant location. Represents project beginning mid-1977, ending mid-1980. Average basis for scaling, mid-1979. Investment requirements for fly ash disposal excluded. Construction labor shortages with overtime pay incentive not considered.

[b] Each item of direct investment includes total equipment costs plus installation labor, and material costs for electrical, piping, ductwork, foundations, structural, instrumentation, insulation, and site preparation.

TABLE A-IV. ESTIMATED AVERAGE ANNUAL REVENUE REQUIREMENTS FOR A 500 MW APPLICATION OF THE HITACHI ZOSEN PROCESS[a]

Item	Annual quantity	Unit cost, $	Annual cost, $	% of annual revenue required
Direct Costs				
Raw materials				
NH_3	5.25×10^6 kg	0.165/kg	866,300	8.47
Catalyst			5,125,000	50.14
Total raw materials			5,991,300	58.61
Conversion costs				
Operating labor and supervision	8,760 labor hrs	12.50/labor hr	109,500	1.07
Utilities				
Steam	86,700 GJ	1.896/GJ	164,400	1.61
Electricity	10,787,000 kWh	0.029/kWh	312,800	3.06
Heat credit	92,400 GJ	-1.896/GJ	(175,200)	(1.71)
Maintenance = (0.04) (TDI)			470,200	4.60
Analyses	2,920 labor hrs	17.00/labor hr	49,600	0.48
Total conversion costs			931,300	9.11
Total direct costs			6,922,600	67.72
Indirect Costs				
Capital charges				
Depreciation = (0.06) (total depreciable investment)			1,287,300	12.59
Average cost of capital = (0.086) (total capital investment)			1,900,300	18.59
Overheads				
Plant = (0.5) (conversion costs minus utilities)			314,700	3.08
Administrative = (0.1) (operating labor costs)			11,000	0.11
Total indirect costs			3,513,300	34.37
Spent catalyst disposal			(214,000)	(2.09)
TOTAL ANNUAL REVENUE REQUIREMENTS			10,221,900	100.00

[a] Basis: 500 MW new coal-fired power plant, 3.5% S coal. 90 percent NO_x reduction, 90 percent SO_2 removal. Midwest power plant location, 1980 revenue requirements. Remaining life of power plant = 30 years. Plant on line 7000 hr/yr. Plant heat rate = 9.5 MJ/kWh. Investment and revenue requirement for disposal of fly ash excluded. Total direct investment $11,754,000; total depreciable investment $21,455,000; and total capital investment $22,096,000.

This Exxon report describes their thermal DeNOx system. It has proven successful on industrial and utility boilers, steamers, furnaces and incinerators primarily in Japan and California. Cost data are provided in figure 1 (p. 485).

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THERMAL DeNO$_x$: THE PRACTICAL APPROACH TO DEEP NO$_x$ REDUCTION

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ABSTRACT

Thermal DeNO$_x$ is a commercially proven process which achieves selective non-catalytic reduction of NO$_x$ in flue gas from stationary combustion sources. This thermally driven process involves the injection of ammonia (NH_3) into high temperature flue gas in the 850C-1200C plus temperature range. Recent development of a fundamental kinetic model of the process chemistry has provided increased insight into process capability and applicability. This has permitted development of improved technology resulting in higher deNO$_x$ performance capability, better load following, more practical injection systems and substantially lower capital requirements.

INTRODUCTION

In the past decade, nitrogen oxides (NO$_x$) have emerged as a serious air pollutant in addition to sulfur oxides (SO$_x$). Both of these pollutants are normally found in flue gas emitted from combustion sources; and, if they exist in sufficient concentration, contribute to respiratory health and visibility problems. More recently, the issue of acid rain has prompted further debate and closer scrutiny of permissible emission levels of these pollutants. This paper will describe the Exxon Thermal DeNO$_x$ process which is a relatively inexpensive process for the deep reduction of NO$_x$ emissions.

In order to achieve significant reductions in the emission of NO_x, some form of control technology is required. Those most commonly used include low-NO_x burners, Thermal $DeNO_x$, and selective catalytic reduction (SCR). Based on data compiled and published by the Air Resources Board of the state of California (CARB) (see reference), the most cost effective technology is low-NO_x burners and the least cost effective is SCR. As illustrated in Fig. 1 for retrofit of these technologies to petrochemical equipment, it can be seen that Thermal $DeNO_x$ now approaches the cost effectiveness of low-NO_x burners and achieves performance approaching that of SCR. By combining low-NO_x burners and Thermal $DeNO_x$, performance equivalent to SCR can be achieved at substantially lower investment.

Thermal $DeNO_x$ involves injection of NH_3 into the hot flue gas in a prescribed temperature range. This temperature range can be significantly expanded through the additional injection of hydrogen. The NO_x reduction is essentially independent of the concentration of oxides of sulfur or particulate matter in the flue gas.

Thermal $DeNO_x$ is a commercially demonstrated process with installations in gas and oil-fired steam boilers, utility boilers and process furnaces. Successful tests have also been conducted on a municipal incinerator, oil field steam generators and a glass melting furnace. Tests on flue gas generated by coal combustion have demonstrated the applicability of the process to coal-fired utility boilers (Varga, et al, 1979), but there is no commercial experience to date.

Recently, development of an accurate kinetic model of the process chemistry has significantly expanded the understanding of the interation of all parameters affecting performance, and permits accurate calculation of performance and ammonia breakthrough. Simultaneously, novel injector designs were field tested and revealed substantially greater performance capability of 70% to over 80%, load following capability without the use of hydrogen, gridless injectors and a lower cost installation. In effect, Thermal DeNO$_x$ now offers performance approaching that of competing flue gas treatment processes for many types of fired equipment, while retaining or improving upon its historical advantages in cost and space requirements.

PROCESS DESCRIPTION

Thermal DeNO$_x$ is a non-catalytic process based on the gas phase homogeneous reaction between the NO$_x$ in flue gas and ammonia (NH_3) which produces nitrogen and water. In general, NH_3 is injected into the hot flue gas by means of either air or steam carrier gas at a point in the flue specifically selected to provide optimum reaction temperature and residence time. The following two equations describe the reactions which govern the overall process:

$$NO + NH_3 + O_2 + H_2O + (H)_2 \rightarrow N_2 + H_2O \qquad (1)$$

$$NH_3 + O_2 + H_2O \rightarrow NO + H_2O \qquad (2)$$

As indicated by equation 1, hydrogen (H_2) can be injected along with NH_3 to extend the effectiveness of the deNO$_x$ reaction down to the range of 700C.

Detailed laboratory experiments have shown that there are no reactions between the Thermal DeNO$_x$ process and sulfur compounds in high temperature flue gas regions. Also, the injection of NH_3 does not cause SO_2 to undergo additional oxidation to SO_3.

In practice, some of the injected NH_3 escapes reaction with the NO$_x$ and, as it cools with the combustion gases, it reacts with SO_3 and H_2O to form NH_4HSO_4, and $(NH_4)_2SO_4$. Laboratory as well as full-scale commercial test results show these sulfates to be harmless with respect to both corrosion and preheater fouling when the Thermal DeNO$_x$ process is operated within the prescribed design specifications. Long-term tests conducted in two oil-fired boilers in Japan showed that these sulfate deposits are readily removed by washing with water at prescribed intervals.

Initial tests on a 0.9 MW combustion system (Varga, et al, 1979) to determine the applicability .of Thermal DeNO$_x$ to coal-burning utility boilers gave results similar to those obtained by natural gas firing in the same system. This study, which was carried out under a joint Exxon/EPRI-sponsored program, represents the initial development step in the application of the process to coal combustion.

KINETIC MODEL DEFINES INTERACTION OF PERFORMANCE PARAMETERS

Equations (1) and (2) summarize the overall effect of the deNO$_x$ reaction, but they do not reveal the complexity of interaction of numerous intermediate chain branching reactions involving certain key free radicals. The chemical mechanism includes the usual H, OH, O and H_2O reactions as well as those shown in Fig. 2.

This set of rate equations was numerically integrated to yield concentration-time profiles which were compared to an extensive set of experimental data on Thermal DeNO$_x$ as well as to data on the kinetically related system of ammonia oxidation. With this fundamental kinetic model, it was possible to achieve good fits ($\pm$7%) to these data using kinetically plausible rate constants. An example of the model calculation versus data is shown in Fig. 3. In general, the observed differences between the calculated and observed profiles are consistent with experimental uncertainty.

The kinetic model has been successful in assessing the process capability through a more precise calculation of the interaction of kinetic performance parameters listed in Table 1. This technology has enabled extension of the temperature window using "NH_3" alone from under 870C to over 1200C under certain conditions of the other kinetic parameters. The model can be used to pinpoint the optimum location in a given piece of equipment for NH_3 injection or to evaluate the optimum amount of NH_3 which should be injected in a given location. The model also permits accurate prediction of unreacted ammonia. An illustration of model calculations is shown in Figure 4 for a utility boiler at two different loads. You will note that this performance is based on the use of wall injectors and these are described in the next section.

ENGINEERING PERFORMANCE PARAMETERS

The kinetic model calculations represent kinetic limit performance for the process and this may be lowered in a real system due to one or more of the engineering performance parameters listed in Table 1. The influence of each of these parameters has been the subject of considerable research and development and some work in this area is being continued. Mixing effectiveness of grid injectors has been increased such that it is possible to achieve 95-100% of the kinetic limit performance (Hurst and Schleckser, 1980). Wall injector mixing is the subject of current R&D activity and field test work.

Selection of the optimum injector location is primarily a function of temperature and residence time. Long residence time (over 1 second) at optimum temperature will tend to give relatively high performance even with non-optimum initial mixing or adverse temperature and velocity gradients. However, mixing and temperature/velocity gradients take on primary importance when the zone of injection involves steep temperature decline, coaxial and transverse to flow direction. Also, velocity gradients at the injector zone make it difficult to achieve uniform NH_3/NO_x ratios throughout the flue gas stream and this results in a loss of process efficiency. The kinetic model is capable of calculating the influence of these factors based on field measurements or studying their effect based on assumed operating conditions. In addition, R&D work is underway to develop analytical techniques for predicting mixing effectiveness of both wall and grid injectors in real flue gas systems.

OTHER TECHNOLOGY ADVANCES

The original Thermal $DeNO_x$ applications involved positioning an array of injector pipes or an injection grid within the flue gas stream at the proper flue gas temperature for injecting a mixture of NH_3 and carrier steam or air. Sometimes, multiple grids and/or hydrogen were employed to achieve some load following, but this increased the complexity and cost of the installation. Also, many potential users voiced concern over the long term operability of the grid particulary in corrosive, erosive or heavy fouling flue gas streams. In most applications today, the injection grids are being superseded by wall injectors which consist of large jets located at or near the boundary walls of the injection zone.

The wall injectors are simple and relatively inexpensive, but require careful design and placement in order to achieve optimum process results. They are not subject to overheating upon loss of carrier, nor do they face corrosive attack and/or collection of fuel ash deposits such as may result from firing heavy fuel oil or coal. These factors permit location of the injectors at the optimum flue gas temperature even within the combustion zone of fired equipment. To accommodate turndown operation, two sets of injectors may be used with very little impact on overall cost. Through proper location of the injectors to cover the load range, the necessity of hydrogen injection to promote low flue gas temperature performance is minimized. In addition, if flue gas temperatures during operation of a unit are substantially different from those anticipated in the design, these type injectors may be relocated either onstream or during a short outage.

Demonstration of this new technology has been performed on a small refinery heater and a steam generator. Performance results for the refinery heater are shown in Fig. 5 and compared with kinetic model calculations. Even though the injectors used in this test were not optimized with regard to number or location, the performance achieved approaches the kinetic limit predictions and demonstrates the superiority of the new technology. The steam generator results, shown in Fig. 6, illustrate an optimized application of this technology where deNO$_x$ exceeding 90% was achieved.

PROVEN PERFORMANCE

Thermal DeNO$_x$ has been successfully demonstrated in gas and oil-fired utility boilers, package and field erected industrial boilers, process furnaces, an oil field steamer, and a glass melting furnace as summarized in Table 2. Even through these installations represent the application of older technology, they provide a firm basis for building the new technology available today. In all of these installations where Thermal DeNO$_x$ has been operated in accordance with its design conditions, performance has been in accordance with predicted values.

Using the new technology outlined above, ~~it is~~ possible to project performance for general types of fired equipment used in various industries as shown in Table 3. While these are general ranges of process capability, it must also be cautioned that the process is equipment specific and should be evaluated for each potential application.

The process has a wide range of applicability ranging from clean flue gas services to high ash, high fouling services. It is an excellent supplement to low NO_x combustion schemes where additional NO_x reduction is required. Applicability and effectiveness will vary from one unit to the next depending on local flue gas conditions and on the configuration of the high temperature zones. Preliminary estimates of performance expected to be achieved for retrofitting Thermal $DeNO_x$ into existing units can be made on the basis of engineering drawings and design specifications. Recommendations can also be provided for designing new boilers and heaters and for modifying existing designs so as to maximize the efficiency of the process without compromising normal design and operating features.

HOW MUCH DOES IT COST?

While it is recognized that, in general, air pollution devices only contribute to a plant's operating expenses and are only installed as the result of legislation, when they do become necessary one of the first questions asked is in regard to cost. The cost of Thermal $DeNO_x$ is below that of other flue gas treatment processes but higher than combustion modifications in most applications. The recent technology advances have both reduced cost and provided increased performance, thus yielding a more cost effective process.

In order to illustrate the cost effectiveness of the Thermal $DeNO_x$ process, a 200,000 lb/hr oil- or gas-fired industrial boiler has been selected for detailed analysis. Figure 7 shows the corresponding flow schematic of the Thermal $DeNO_x$ supply system, which is sized on the basis of the values presented in Table 4. It is assumed that the boiler has an uncontrolled NO_x effluent of 200 vppm corrected to 3% O_2 dry and that it will be equipped with two sets of injectors for injecting NH_3 via a steam carrier. Provision for a one-month NH_3 supply storage facility is also assumed in this analysis.

The total erected capital investment for these facilities is estimated to be $389,000 as outlined in Table 5. Direct costs include the material and labor for equipment outlined in Table 4 plus interconnecting piping. Indirect costs include field labor overheads, construction supervision and equipment, labor wage taxes, erection fee, engineering costs, and licensing fee. Contingency is based on 30% of direct, plus indirect cost less licensing fee. Costs are expressed at 2Q82 Southern California level. Excluded from these costs are such items as land and owners' charges.

Cost effectiveness for the sample boiler installation is also shown in Table 5 for three different bases. Annualized costs are obtained by assuming a five-year payout on investment plus annual operating costs. On a heat-fired basis, cost effectiveness of the process is $0.38/MW. The cost effectiveness based on NO_x removed is $0.23/kg NO_2 for a $DeNO_x$ efficiency of 70% and $0.20/kg NO_2 for 80% $DeNO_x$. This cost compares favorably with some combustion modification techniques, such as low-No_x burners, and is substantially lower than selective catalytic reduction.

Annual operating costs are shown in Table 6. These costs are based on a 65% annual load factor and include values for NH_3, power and steam consumption, and maintenance material and labor.

Similar costs can be constructed for specific applications on most any type of stationary combustion equipment. Table 7 shows the approximate investment cost for various industrial applications. In each case it is assumed that the process can be applied with wall injectors; and, in every case except single checker glass furnace, no H_2 is required. In considering these estimates of cost, it must be cautioned that costs with this process are site and equipment specific and costs could vary substantially for a specific application.

SUMMARY

In summary, the Thermal DeNO$_x$ process offers one of the most practical and cost-effective approaches to significant NO$_x$ reduction from stationary fired equipment of currently available post-combustion NO$_x$ removal processes. It is a proven process, applicable to most types of fired equipment. Recent technological advances, based on a fundamental understanding of the process chemistry, have resulted in significant increases in Thermal DeNO$_x$ performance along with decreases in equipment cost. As a result, Thermal DeNO$_x$ now provides cost effectiveness approaching that of low-NO$_x$ burners while achieving deNO$_x$ performance approaching that of SCR.

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Figure 1

COST EFFECTIVENESS FOR RETROFITTING VARIOUS NO_X CONTROL TECHNOLOGIES TO PETROCHEMICAL COMBUSTION EQUIPMENT

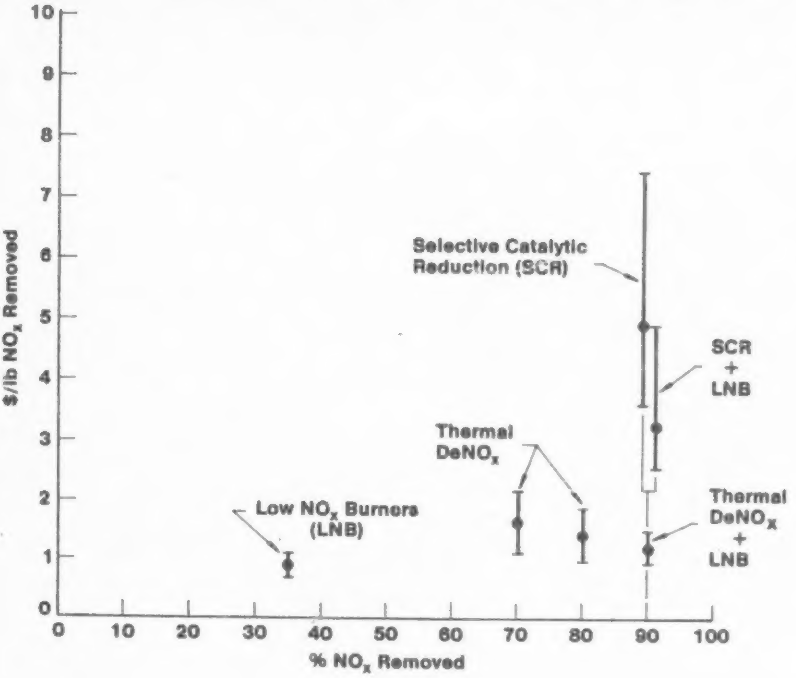


Figure 2

MAIN REACTION PATHS FOR NITROGEN CHEMISTRY

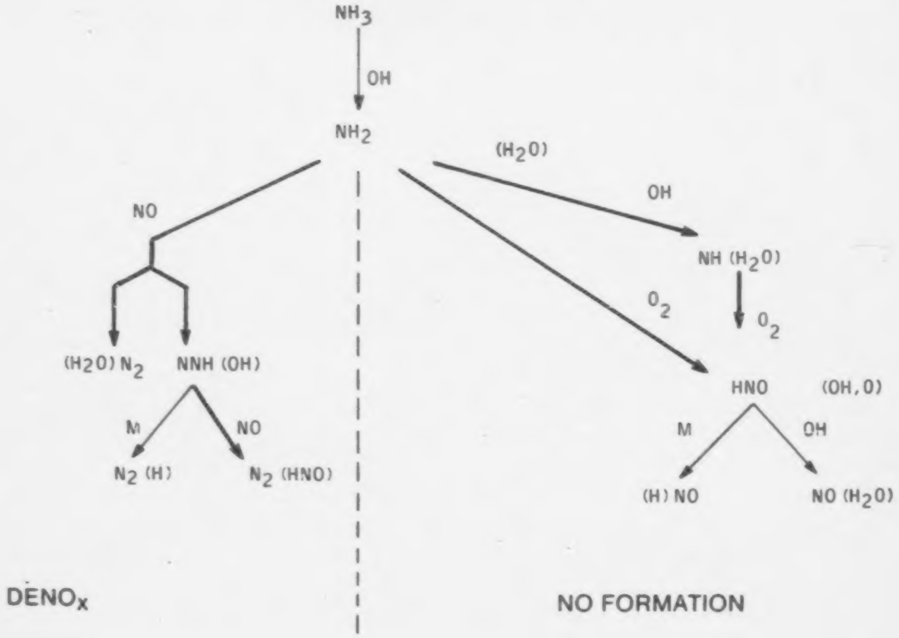

NH3
OH
NH2
(H2O)
OH
NO
NH (H2O)
O2
O2
(H2O) N2
NNH (OH)
HNO
(OH, O)
M
NO
M
OH
N2 (H)
N2 (HNO)
(H) NO
NO (H2O)
DENOx
NO FORMATION

Figure 3

KINETIC MODEL PREDICTION vs. LABORATORY TEST DATA

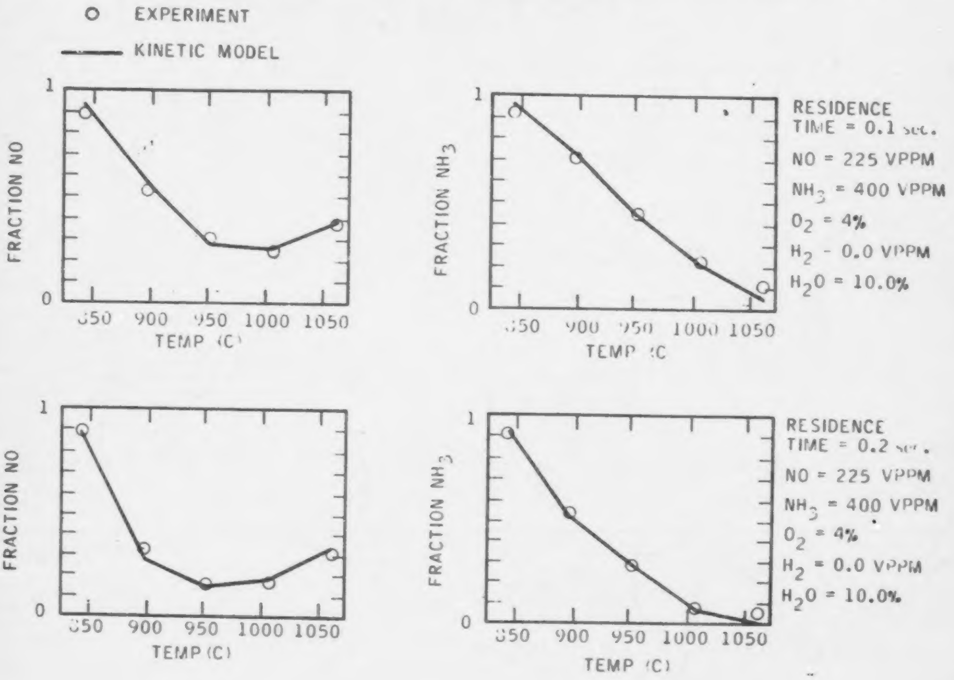


Figure 4

DeNO$_x$ AND UNREACTED NH$_3$ PREDICTION AT 100% AND 50% LOAD

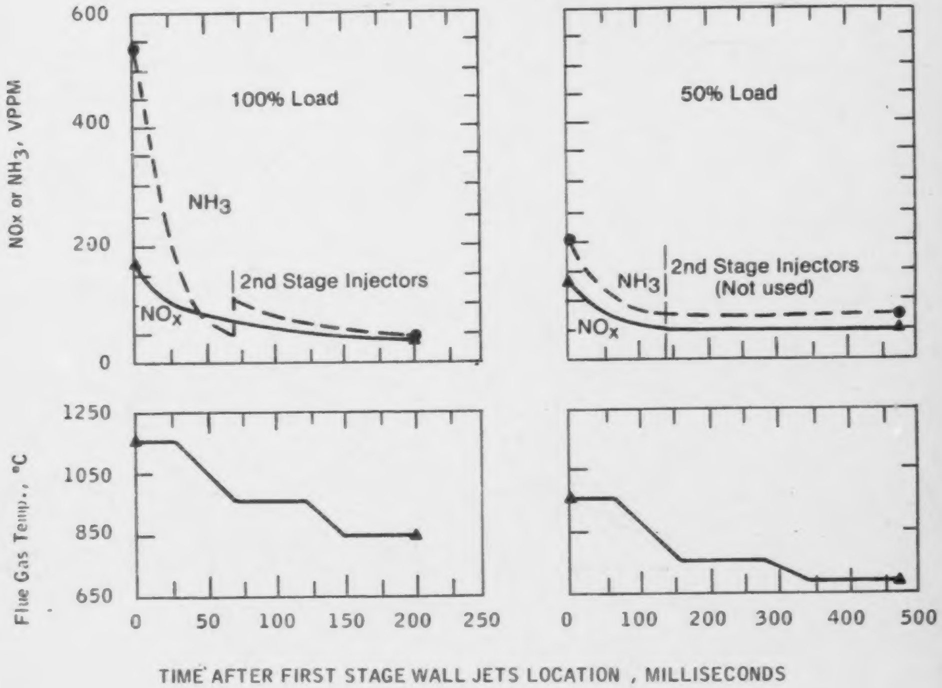
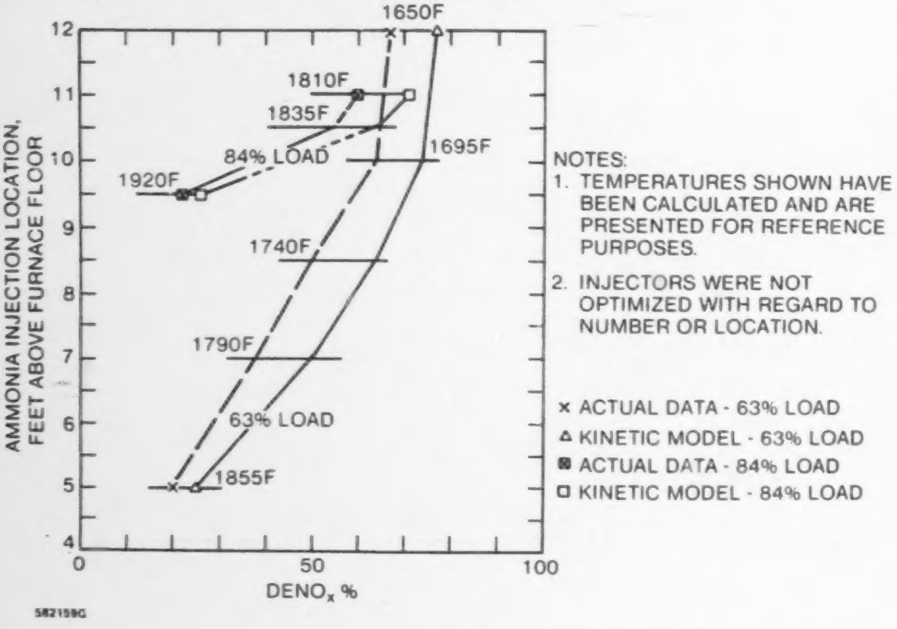


Figure 5

KINETIC MODEL PREDICTION VS ACTUAL DATA FOR A TYPICAL REFINERY UNIT



NOTES:

1. TEMPERATURES SHOWN HAVE BEEN CALCULATED AND ARE PRESENTED FOR REFERENCE PURPOSES.
2. INJECTORS WERE NOT OPTIMIZED WITH REGARD TO NUMBER OR LOCATION.

Figure 6

DATA FROM THERMAL DENO$_X$ DEMONSTRATION TEST ON AN OIL FIELD STEAM GENERATOR

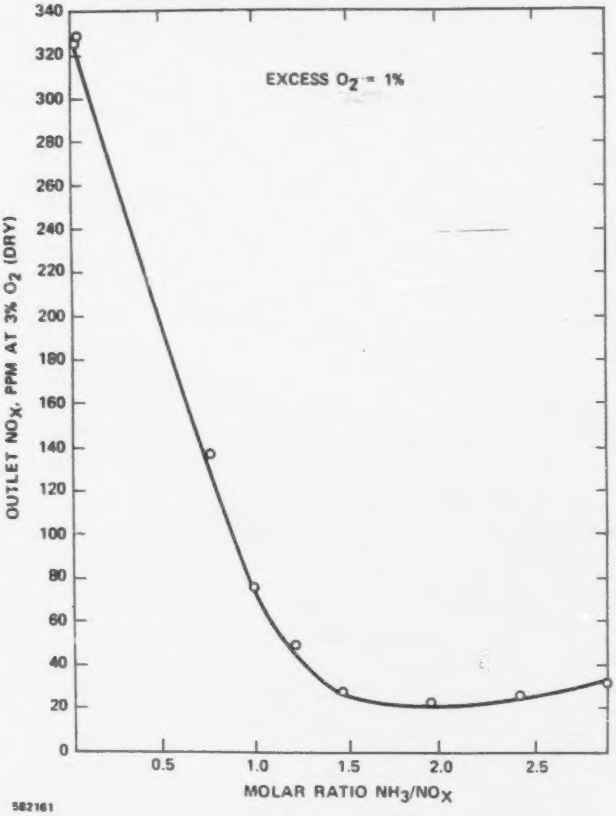


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Figure 7

SIMPLIFIED THERMAL DeNO$_x$ SUPPLY SYSTEM FLOW DIAGRAM

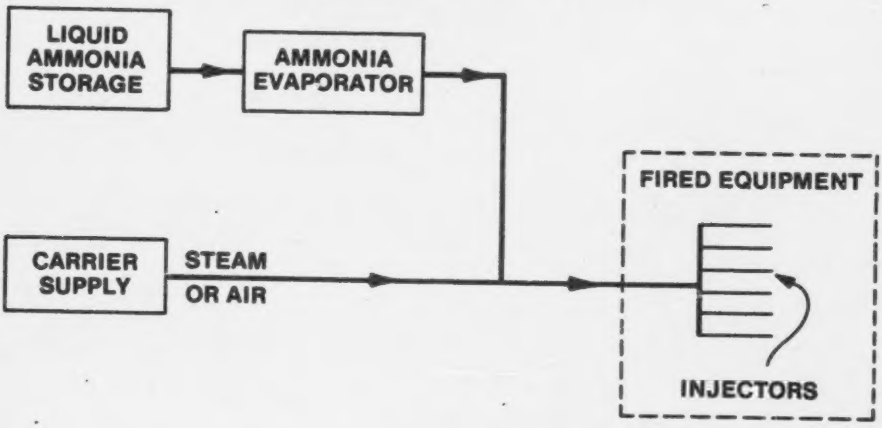
LIQUID AMMONIA STORAGE
AMMONIA EVAPORATOR
FIRED EQUIPMENT
CARRIER SUPPLY
STEAM
OR AIR
INJECTORS

Table 1

THERMAL DeNO$_x$ PERFORMANCE PARAMETERS

KINETIC	ENGINEERING
Temperature	Mixing
Time	Carrier Rate
Initial NO$_x$	Temperature Gradient
NH$_3$/NO$_x$	Velocity Gradient
O$_2$	Staging
H$_2$O	
Equilibrium level of combustion free radicals such as H, OH, etc.	

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Table 2

THERMAL $DeNO_x$ EXPERIENCE SUMMARY (OIL AND GAS FIRED UNITS)

	OPERATIONAL	CONSTRUCTION PHASE	DESIGN PHASE
Japanese Industrial Boilers	3		
Japanese Utility Boilers	5		
Japanese Petroleum Heaters	4		
California Oil Field Steamer (Demonstration)	1		
California Flat Glass Melting Furnace	1		
California Petroleum Heaters	19	5	2
California Utility Boiler	1		
California Industrial Boilers	3		
California Incinerators	1		1
California Oil Field Steamers (Commercial)	3		

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Table 3

TYPICAL THERMAL DeNO$_x$ PERFORMANCE WITH IMPROVED TECHNOLOGY

	% DeNO$_x$
Petroleum Heaters	75-85
Industrial Boilers	
— Package	50-70
— Field Erected	70-80
Utility Boilers	60-80
Incinerators	70-80
Glass Furnaces	50-80
Cement Kilns (Pre-calciner)	70-90

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Table 4
EQUIPMENT SIZING BASIS FOR PERMANENT THERMAL DeNO$_x$ FACILITIES FOR SAMPLE INDUSTRIAL BOILER

BOILER DESIGN CONDITIONS 91t/h, 4.2 MPag, 370C	
Fuel	Oil or Gas
Initial NO$_x$	200 vppm
NH$_3$/NO$_x$	2.0
CARRIER Requirement	Steam 0.19 kg/s 0.1 MPag (Minimum)
NH$_3$ Storage Tank	30 Day Storage Capacity (38 m^3)
NH$_3$ Vaporizor	(1) Electric Element, Direct Contact
Injectors	Two Locations
Instrumentation	As Required

Table 5
THERMAL DeNO$_x$ INVESTMENT COST AND COST FFECTIVENESS FOR SAMPLE INDUSTRIAL BOILER

Capital Investment(1)	
Material and Labor(2)	$108k
Indirect Costs(3)	$191k
Contingency(4)	$ 90k
	$389k
Cost Effectiveness	
$/MW Fired	0.38
$/kg NO$_x$ Removed @ 70% DeNO$_x$	0.23
$/kg NO$_x$ Removed @ 80% DeNO$_x$	0.20

Notes:
(1) Investment costs are expressed at 2Q82 Southern California level
(2) Includes equipment items listed in Table 4 plus interconnecting piping
(3) Includes field labor overheads, construction supervision and equipment, labor wage taxes, erection fee, contractor engineering and Exxon Research and Engineering Co. charges. Excluded are such items as land, owners charges and licensing fee.
(4) Contingency fee is based on 30% of direct costs plus indirect costs.

Table 6
THERMAL DeNO$_X$ ANNUAL OPERATING COST FOR SAMPLE INDUSTRIAL BOILER

Item	Annual Consumption(1)	Unit Cost	Annual Cost
Ammonia(2)	164 t	$220/t	$36,080
Electric Power(3)	55 MWh	$50/MWh	2,750
Steam(4)	3890 t	$13.50/t	52,515
Maintenance			
Material and Labor(5)			7,560
		Total	$98,905

Notes:

(1) Assumes 65% load factor
(2) Includes ammonia for direct injection
(3) Includes power requirement for ammonia vaporizer
(4) Low pressure steam (0.1 MPag minimum) for carrier
(5) Assumed to be 7% of direct investment cost

Table 7

TYPICAL THERMAL DeNO$_x$ COST WITH IMPROVED TECHNOLOGY

	Investment $/kWt
Petroleum Heaters	3-9
Industrial Boilers — Package — Field Erected	4-9
Utility Boilers	5-10
Incinerators	6-9
Glass Furnaces	6-11
Cement Kilns (Pre-calciner)	3-6

This EPRI paper describes in some detail combustion modification approaches to NOx control and dry sorbent injection processes for SO2 control. Each of these is a relatively low-cost, retrofitable pollution control method; combination of the two concepts may offer a particularly cost effective approach for simulataneous NOx and SO2 removal. Pollutant removal efficiencies, cost and reliability are also discussed for each technological approach.

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MAY 1983

RETROFIT NO_x AND SO_2 CONTROLS FOR COAL-FIRED UTILITY BOILERS

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ABSTRACT

Regulatory pressures for retrofit control of SO_2 and NO_x are mounting due to increased national awareness of NO_x and SO_2 emissions, revised plans by utilities regarding retirement of existing units, and new environmental concerns such as acid pricipitation. Historically, NO_x control has been achieved cost effectively through combustion modification. Combustion modifications continue to be the preferred NO_x control approach from the standpoint of NO_x removal effectiveness, cost, and reliability compared to flue gas treatment.

Dry sorbent injection processes, either in-furnace or post-furnace, are potentially attractive SO_2 removal processes because of their simplicity and cost effectiveness. Renewed interest in in-furnace sorbent injection for SO_2 control is motivated by recent pilot-scale tests which indicate that the thermal environment produced by low-NO_x combustion systems may optimize the in-furnace SO_2 capture process providing combined in-furnace NO_x/SO_2 control. Post-furnace injection currently appears to be most applicable to units firing low sulfur western coals.

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SECTION 1.0

INTRODUCTION

Emissions of NO_x and SO_2 have been regulated for new utility boilers since the passage of the federal Clean Air Act and the establishment of New Source Performance Standards (NSPS) in 1971. For units designed prior to 1971 (hereafter referred to as "pre-NSPS"), there are no federal SO_2 or NO_x emission standards. However, regulatory pressure to control these emissions is mounting due to increased national awareness of NO_x and SO_2 emissions, revised plans by utilities to slow down the retirement of older units, and new environmental concerns such as acidic precipitation.

Pre-NSPS utility boilers, particularly those which are coal-fired, are a primary target for new emission standards because of their relative contribution to national SO_2 and NO_x levels. As illustrated in Figure 1-1, electric utility boilers represent 65 percent and 29 percent of the total SO_2 and NO_x emissions, respectively. Coal is the most common fuel for reasons of price and availability and it dominates the utility emissions as shown in Figure 1-2. The utility industry may also expect continued regulatory pressure because (1) the emissions are concentrated at a stationary point source considered by some to be easier to control, and (2) the cost effectiveness of controlling stationary source emissions is perceived to be greater than that for the transportation sector.

The number of coal-fired power plants built since 1971, and to be constructed in the near future, is significantly below projections made in 1971. Utilities have and will continue to rely on pre-NSPS units for a major portion of generating capacity. Consequently, regulatory attention for existing sources is intensifying.

1.1 RATIONALE FOR IN-FURNACE NO_x/SO_2 CONTROL

Historically, NO_x control has been more cost effectively achieved by means of minimizing formation in the furnace section through combustion

DISTRIBUTION OF NO_x AND SO_x EMISSIONS BY SOURCE

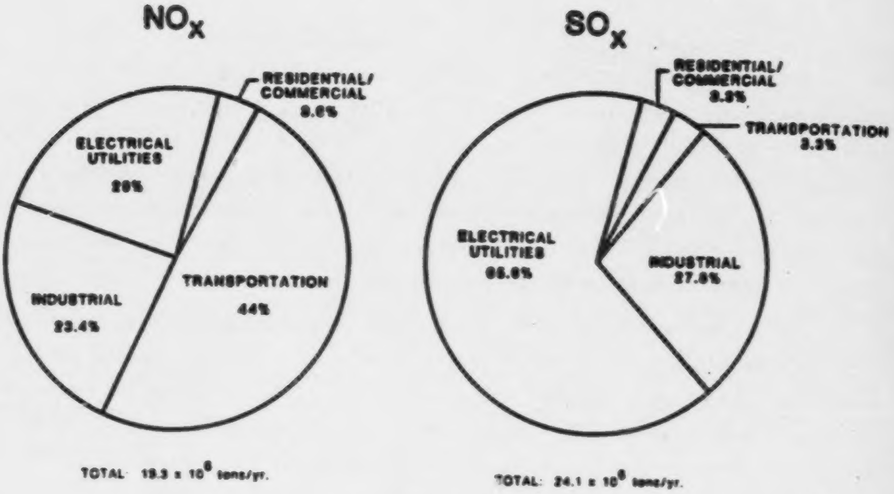


Figure 1-1.

UTILITY NO_x AND SO_x EMISSIONS BY FUEL

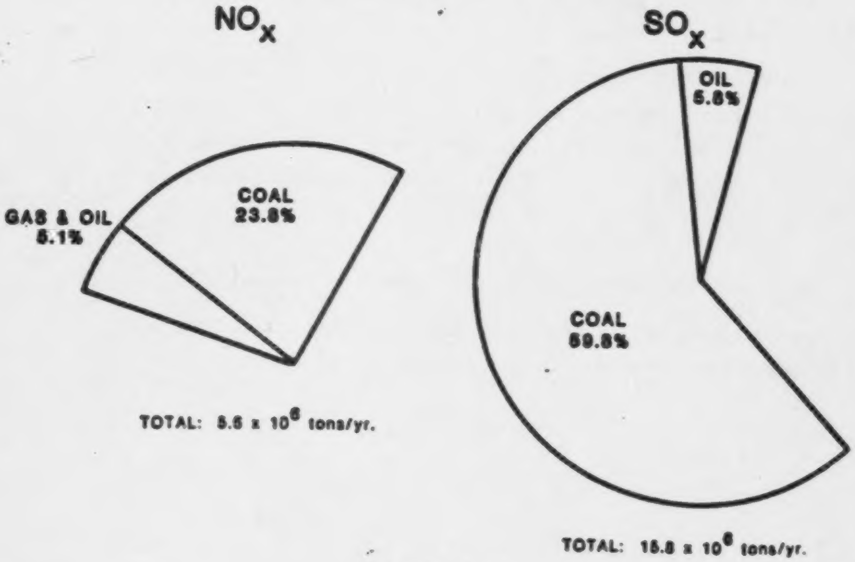


Figure 1-2.

modifications. More than 69 coal-fired units are currently operating in accordance with NSPS regulations. Combustion modifications applied in the form of low-NO_x burners are the basis for NO_x control at coal-fired power plants designed after 1971. Because of their relative simplicity, combustion modifications have had little or no impact on the operation and reliability of new plants and their incremental costs have been minimal (generally less than $5/kW).

The availability of retrofit NO_x combustion controls has become an especially significant issue with the recent emergence in Japan of post-combustion flue gas treatment processes for retrofit NO_x control. There are presently over 9000 megawatts of flue gas treatment NO_x controls installed and operating at over 30 oil- and coal-fired Japanese utility boilers, and most of these systems have been retrofitted. Although these processes offer the prospect for substantial reductions in NO_x emissions below what is now possible with combustion controls, their capital cost is estimated to be 10 times that of combustion control--and unlike combustion controls, their operating costs are substantial. For reasons of cost effectiveness and reliability, combustion controls offering significant NO_x reductions are the preferred NO_x control approach.

Until recently, the control of SO_2 emissions was primarily achieved through either flue gas treatment (e.g., wet scrubbers) or through changes in fuel sulfur content (e.g., coal cleaning or low-sulfur coal). Utilities have long sought a cost-effective alternative to wet scrubbers for SO_2 control and the recent successes with spray-dry and dry injection sorbent processes have led to increased interest in exploring direct SO_2 control in the furnace section of the boiler.

Figure 1-3 illustrates the various approaches to SO_2 removal. A progression from a wet scrubber to a spray dryer system provides a moderate simplification to the hardware requirements, slightly reduces water usage, produces a dry material for disposal and provides a small cost savings over wet scrubbing. Further simplification and cost savings can be accomplished by utilizing an all dry SO_2 removal system. It should be noted, however, that

HOLD

SO_2 REMOVAL APPROACHES

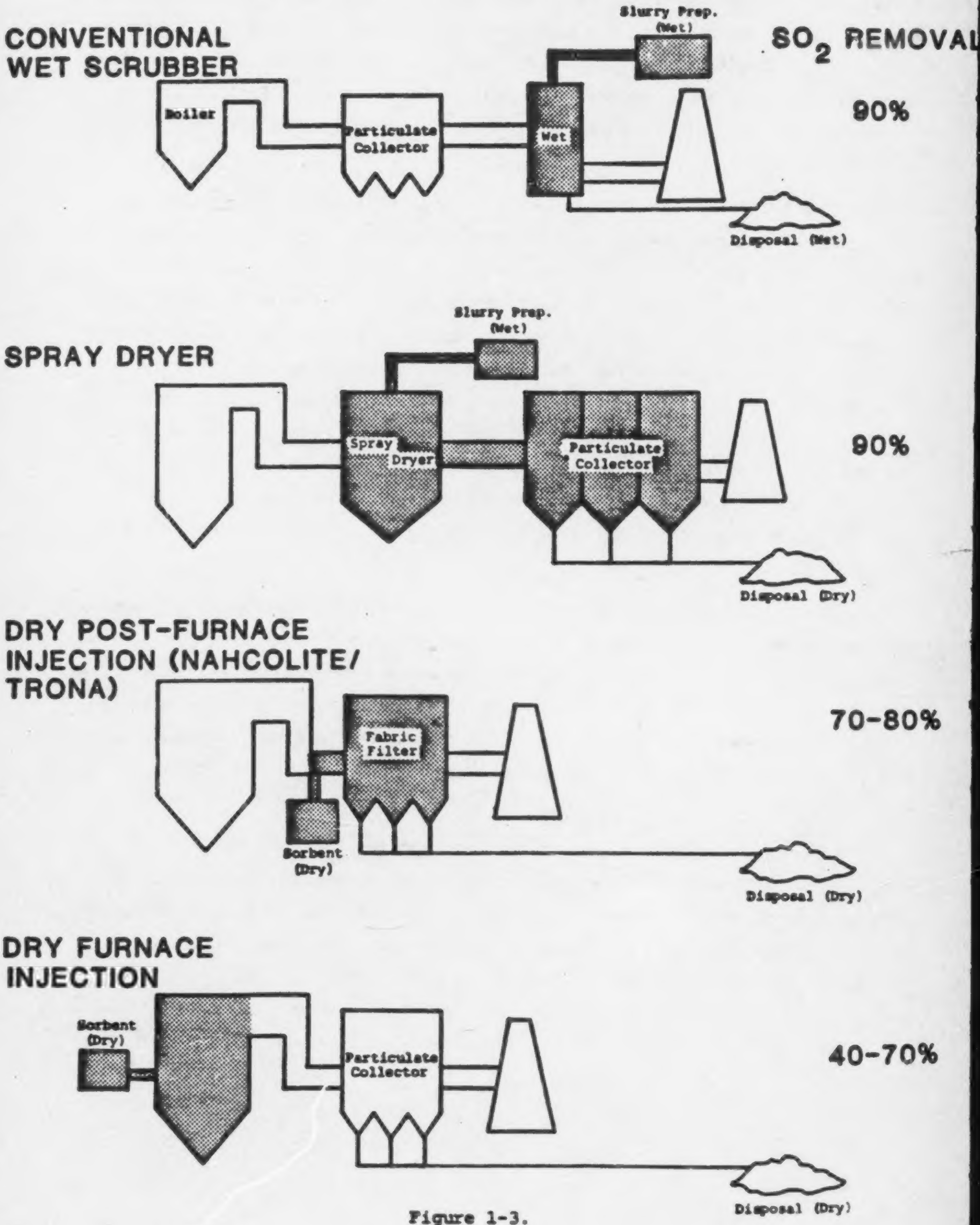

CONVENTIONAL WET SCRUBBER
Slurry Prep. (Wet)
SO_2 REMOVAL
Boiler
Particulate Collector
Wet
90%
Disposal (Wet)
SPRAY DRYER
Slurry Prep. (Wet)
Spray Dryer
Particulate Collector
90%
Disposal (Dry)
DRY POST-FURNACE INJECTION (NAHCOLITE/ TRONA)
Fabric Filter
Sorbent (Dry)
70-80%
Disposal (Dry)
DRY FURNACE INJECTION
Sorbent (Dry)
Particulate Collector
40-70%
Disposal (Dry)

Figure 1-3.

although the all dry processes appear simpler and cheaper they are not as effective as the wet processes for SO_2 removal capability.

The two all dry SO_2 removal systems shown in Figure 1-3 are currently under development and will be the subject of Section 2.0 of this document. The first is categorized as a flue gas treatment system and involves injecting a pulverized sodium compound ahead of a fabric filter. The sodium material collects on the bags where it reacts to form solid sodium sulfate. EPRI has successfully demonstrated this technique on a 22 MW full-scale pulverized coal-fired boiler. The second all dry SO_2 removal approach eliminates the need for a sodium-based sorbent and consists of injecting a pulverized calcium-based material directly into the furnace cavity of the boiler. Upon entering the furnace, the sorbent material reacts with SO_2 to form calcium sulfate which is then removed along with the fly ash in the electrostatic precipitator or fabric filter. An overview of the sodium-based dry post-furnace technology will conclude the SO_2 control section.

It should be noted that an important driving force in U.S. development work is the foreign technology contributions by W. Germany and Japan. The direct injection of limestone in a coal-fired utility boiler in W. Germany has led to SO_2 removal levels approaching 70 percent with low sulfur coal. The Japanese have developed several low-NO_x burners, and innovative combustion modification methods producing NO_x levels less than 200 ppm. There are strong incentives to combine these low-NO_x combustion modification techniques with emerging in-furnace SO_2 control measures to achieve combined NO_x-SO_2 removal with cost-effectiveness benefits substantially greater than flue gas treatment processes. EPRI currently views in-furnace NO_x and SO_2 removal processes to be potentially compatible and the justifications for their development is similar as will be outlined in later sections of this summary report. The development of low cost integrated NO_x and SO_2 emission controls may lead to low cost emission controls for new and existing plants.

1.2 EPRI IN-FURNACE NO_x AND SO_2 RESEARCH OBJECTIVES

Emission control processes developed for new utility boilers may or may not be retrofittable to existing units. In particular, a number of special design and engineering requirements must be met for a variety of existing boiler types and burner configurations before a NO_x combustion technique can be applied. Techniques developed for new units have provided the basis for retrofit developments but in some cases (e.g., B&W cyclone boilers and cell-burner units), these are not directly adaptable. At this time only 11 coal-fired utility boilers have been retrofitted in the U.S. with low-NO_x combustion controls and therefore industry-wide cost and technical feasibility for retrofit NO_x controls is uncertain. The EPRI program addresses the critical information and technology needs of the electric utility industry. In particular, the EPRI program will provide: (1) current cost and performance data on commercially available NO_x retrofit options; (2) boiler NO_x retrofit potential dependence on boiler size, age, firing type, manufacturer, and coal type; and (3) national emissions reduction potential. In addition, EPRI is working with several utility boiler manufacturers to test, evaluate, and develop new low-NO_x combustion technology for retrofit applications.

In contrast to the very site-specific design concerns normally associated with NO_x control technology, it appears that in-furnace SO_2 control technology may be more generally applied to different boiler types. There are several research and manufacturing organizations in the U.S. and abroad who are presently working to commercialize this technology for retrofit application. Despite these multiple activities, there remains a critical need to develop better understanding of the process fundamentals and clearer definition of the side effects on boiler operation and performance. Accordingly, the EPRI program will address these critical issues to provide a technical basis for the general application of in-furnace dry SO_2 control technology, low-NO_x combustion systems, and promising integrated systems. Sections 2.0 and 3.0 of this document will present a detailed discussion of the NO_x and SO_2 processes, respectively, and their prospects for application to utility boilers. Section 4.0 will summarize conclusions and recommendations based on these discussions.

SECTION 2.0

COMBUSTION NO_x CONTROL

NO_x emissions from coal-fired utility boilers are directly influenced by the combustion process whereas SO_2 emissions are directly related to the fuel sulfur content. Although the nitrogen content of the coal has some influence, it is not the determining factor. To illustrate this point, two boilers with substantially different combustion processes, but burning the same coal, could have NO_x emissions as different as 400 and 1200 ppm; a three to one ratio. NO_x formation and emissions control is a complex process that is primarily dependent upon boiler and burner design, boiler firing practice, and to a lesser degree, coal type.

The objective of this section is to outline the important parameters in NO_x formation, control options, and factors that influence the selection of a control method. Although much of the NO_x control technology currently in use was developed for new units, additional criteria must be satisifed for retrofit applications. Utilities have expressed concern over the lack of (1) performance and cost information on retrofit options, (2) available retrofit options for certain classes of high NO_x emissions boilers, and (3) guidelines to assist the utilities in addressing major NO_x retrofit issues.

EPRI has organized several programs to directly address these needs as discussed in more detail at the conclusion of the section. Preliminary results from these studies indicate that promising NO_x retrofit options with costs as low as $5/kW will be available for many coal-fired units in the next 3-5 years.

2.1 NO_x FORMATION AND BOILER DESIGN DEPENDENCE

Prior R&D studies have shown that low excess air or fuel-rich combustion zones in the lower furnace region inhibit NO_x formation by (1) minimizing the peak combustion temperature and (2) restricting the amount of oxygen available to combine with nitrogen compounds present in the fuel and air. For coal-fired utility boilers, the latter approach is the primary method of achieving NO_x control. However, one of the primary challenges in burning pulverized coal is to establish the proper mixing, chemistry and temperature conditions to minimize NO_x but still achieve complete combustion. This must be accomplished in a manner that minimizes smoke and carbon monoxide formation, provides safe stable combustion, and does not slag or foul the boiler with ash deposits that inhibit heat transfer.

A better appreciation for coal-fired utility boiler NO_x emissions and their dependence on design and operating parameters can be obtained by reviewing the emission characteristics of typical boilers now in operation. NO_x emissions from coal-fired utility boilers are strongly dependent upon the unit size and firing configuration as shown in Figure 2-1 for pre-NSPS units. Boilers having a wet bottom wall-fired furnace design or cyclone combustors tend to have very high NO_x emissions. Cyclone units burn coal in a high intensity swirling combustion chamber that encourages NO_x formation. For this reason, they were discontinued when the 1971 NSPS regulations went into effect. For comparison purposes, the NSPS NO_x limitations are included in Figure 2-1. Dry bottom single wall-fired units and horizontally opposed units have burner and furnace designs that lead to more moderate NO_x emissions. However, the larger units tend to have proportionally smaller furnace sections and more intense combustion that accounts for the higher NO_x emissions. Tangentially fired boilers are inherently low-NO_x emitters because of their long distributed combustion flame zones.

An important consideration in the development of retrofit NO_x combustion controls is the relative contribution to the total NO_x emissions based on installed capacity of each of the boiler design configurations. In particular, it is important to not only identify units that represent a significant fraction of the total NO_x emissions, but also those whose emissions are disproportionate to their capacity. Figure 2-2 shows that tangentially fired units have the largest fraction of installed capacity for pre-NSPS units greater than 100 MW but that the emissions are proportionately less. In contrast, wall-fired units represent a slightly larger fraction of the total

FULL LOAD NO_x EMISSIONS FOR PRE-NSPS COAL -FIRED UTILITY BOILER CONFIGURATIONS

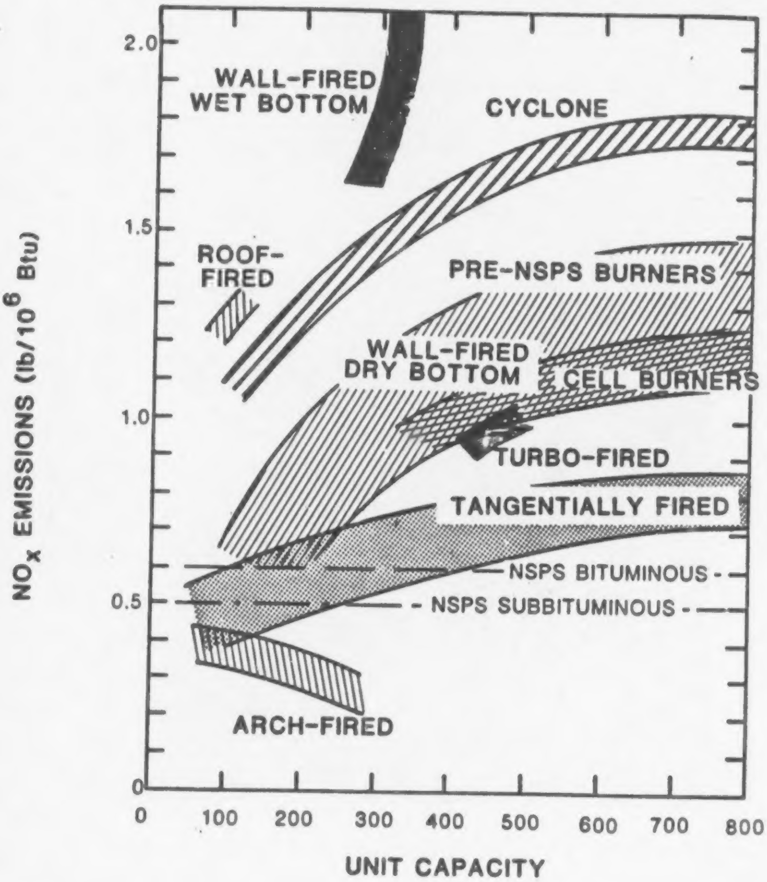


Figure 2-1.

COMPARISON OF TOTAL NO_x EMISSIONS AND UNIT CAPACITY

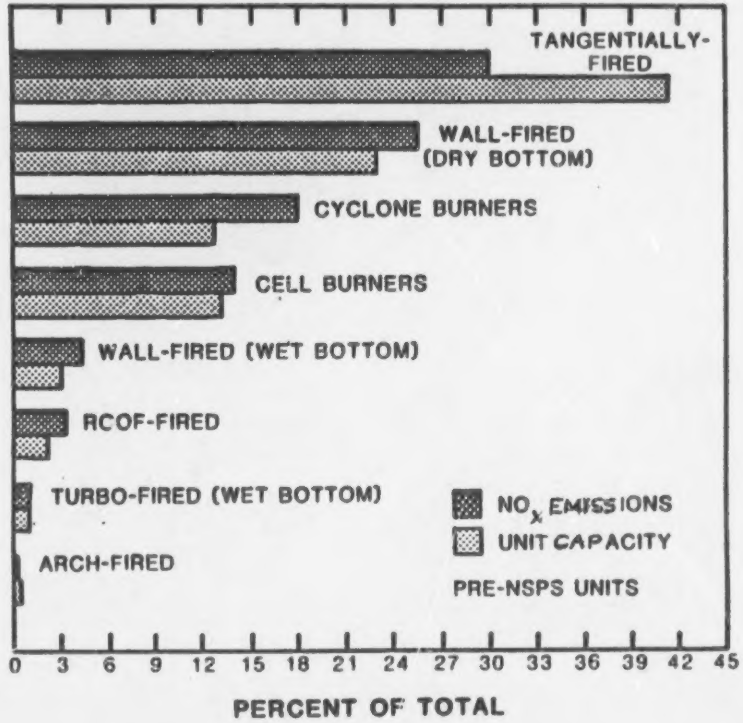


Figure 2-2.

NO_x emissions proportionate to their capacity and the cyclone units are the worst. Another high emitter category is the wall-fired wet bottom boilers but they are not a major concern on a national basis because they represent less than five percent of the total emissions.

2.2 NO_x CONTROL OPTIONS, CONSTRAINTS AND SELECTION CRITERIA

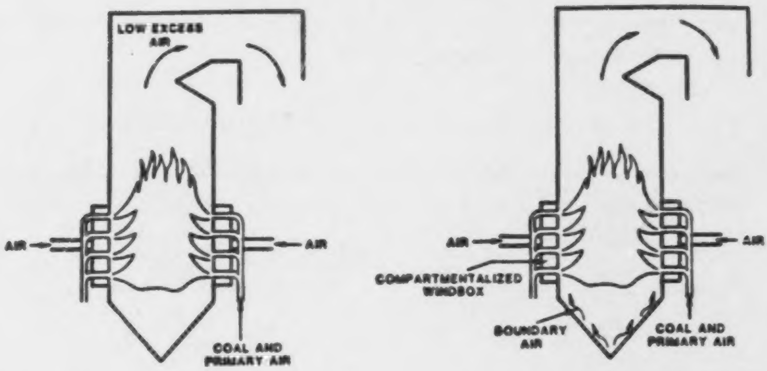
There are basically eleven combustion modification techniques for NO_x control that have been considered for gas-, oil-, and coal-fired utility boilers. They include:

- low excess air combustion
- overfire air or NO_x ports
- biased firing
- burners out-of-service
- flue gas recirculation
- reduced air preheat
- water injection
- low-NO_x burners
- reburning or secondary overfiring
- fuel switching or dual fuel/split fuel
- substantial furnace geometry and design modifications

Flue gas recirculation, reduced air preheat, and water injection are relatively ineffective on coal-fired units. Reductions of less than 15 percent with associated energy penalties and operational concerns make them difficult to justify. Major boiler modifications to replace the lower furnace section with a new firing configuration or burner spacing are very expensive and equally unattactive. From a generic standpoint, the most viable near-term cost-effective combustion control options are low excess air firing, biased firing, overfire air and low-NO_x burners. Schematic illustrations of these options are shown in Figure 2-3.

It should be noted that combustion control options are frequently classified into two major categories of (1) operational modifications, or (2) hardware modifications. Low excess air (LEA) and biased firing are two of the simplest means of achieving NO_x control by operational modifications. Reductions of up to 15 percent (Table 2-1) are not as large as can be achieved with

NOx COMBUSTION CONTROL OPTIONS

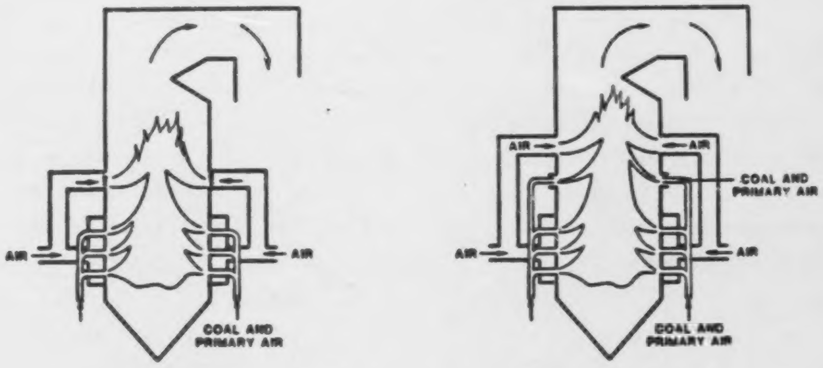

LOW EXCESS AIR
AIR
AIR
COAL AND PRIMARY AIR
AIR
AIR
COMPARTMENTALIZED WINDBOX
BOUNDARY AIR
COAL AND PRIMARY AIR

OPERATIONAL MODIFICATIONS

- LOW EXCESS AIR
- BIASED FIRING

LOW-NOx BURNERS

- CE - PM BURNER
- B&W - DUAL REGISTER
- FW - CF/SF
- RS - CCV


AIR
AIR
COAL AND PRIMARY AIR
AIR
AIR
COAL AND PRIMARY AIR
AIR
AIR
COAL AND PRIMARY AIR

AIR STAGING

- OVERFIRE AIR
- "DEEP STAGING"

FUEL STAGING

- REBURNING
- IN-FURNACE NOx DESTRUCTION
- "MACT"

Figure 2-3.

TABLE 2-1.

COMBUSTION MODIFICATION ASSESSMENT SUMMARY

	CONTROL METHOD									
	LOW EXCESS AIR		BIASED FIRING		BURNERS OUT OF SERVICE		OVERFIRE AIR		LOW NOx BURNERS	
NOx REDUCTION POTENTIAL	TYP.	MAX.	TYP.	MAX.	TYP.	MAX.	TYP.	MAX.	TYP.	MAX.
SINGLE FACE FIRED	0–15%	15%	5%	7%	25%	35%	15%	30%	40%	60%
HORIZONTALLY OPPOSED FIRED	0–15%	15%	5%	8%	25%	35%	15%	50%	40%	60%
TANGENTIALLY FIRED	0–10%	10%	—	--	20%	45%	25%	35%	50%	70%
TOTAL CAPITAL REQUIREMENT ($/kW)	NEW/RETROFIT		NEW/RETROFIT		NEW/RETROFIT		NEW/RETROFIT		NEW/RETROFIT	
SINGLE FACE FIRED	~0/0.40		—		0.20/0.55		1.10/1.80		2.50/7.50	
HORIZONTALLY[(1)] OPPOSED FIRED	~0/0.40		—		0.20/0.55		1.10/1.80		2.50/7.50	
TANGENTIALLY FIRED	~0/0.40		—		0.20/0.55		1.10/1.80		7.9/14.30	

(1) FOR B&W UNIVERSAL PRESSURE UNITS, INCREASE NEW BURNER COSTS BY $5.00/kW, CANNOT OVERFIRE AIR.

hardware modifications such as overfire air ports or low NO_x burners but the implementation cost is minimal. The degree of low excess air operation that can be implemented is dependent upon the margin of excess air above the smoke threshold that was customarily used by the plant operators. If this margin was large, substantial NO_x reductions can be achieved with an associated increase in boiler efficiency. Biased firing achieves a staged combustion effect by biasing coal feeders to achieve more fuel-rich combustion in the lower furnace region and more air-rich in upper elevations. Reductions in NO_x of more than 8-10 percent are difficult to achieve because of coal mill capacity and controls limitations.

Low-NO_x burners are straightforward but the more costly retrofit approach. Consideration must be given to potential physical incompatibility between the retrofit unit and the burner design, however, NO_x reductions can be up to 60-70 percent (Table 2-1). Not all burner retrofits require major modifications. Riley-Stoker has developed a controlled combustion venturi (CCV) burner retrofit that usually does not require complete burner replacement and is relatively inexpensive. Only two field modifications have been completed, but NO_x reductions averaging 50 percent were achieved. CE, B&W, and Foster Wheeler have developed low-NO_x burners with simliar performance that generally require complete replacement of the old burners. No burner replacements are commercially available for B&W cyclone units or cell type burners. Figure 2-4 shows a diagram of the cell type burner which is characterized by a compact burner zone producing high combustion intensity and heat release rates.

A higher degree of staging can be achieved by the installation of overfire air ports as shown in the second portion of Figure 2-3. Overfire air (OFA) achieves a fuel-rich staged combustion condition by diverting combustion air that could normally pass through the burners to ports in the furnace wall above the top row of burners. NO_x reductions of 30 percent or greater can be achieved (Table 2-1). The primary considerations in OFA implementation are (1) cost; (2) proper placement of the ports above the burners to achieve necessary mixing; and (3) lost generation while port holes are cut in the boiler wall. The primary operating constraints are the same as LEA firing

CELL BURNER CONFIGURATION

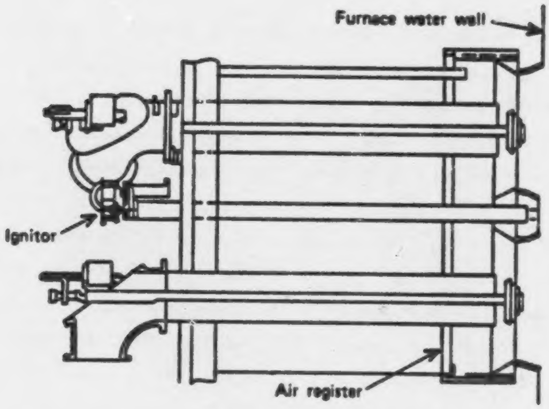

Furnace water wall
Ignitor
Air register

Figure 2-4.

(i.e., smoke, CO, furnace deposits and carbon utilization). Typically, 15 to 20 percent of the combustion air is diverted through the overfire air ports although the Japanese have achieved a "deep staging" condition by diverting up to 30 percent of the combustion air to upper furnace elevations.

Advanced NO_x combustion control concepts are being developed to more carefully control the fuel/air mixing, temperature, and combustion chemistry in the furnace. Alternate fuel-rich/air-rich burner configurations and fuel staging concepts have been developed to separate and control the various combustion zones as shown in Figure 2-3. Much of this development work has taken place in Japan. Mitsubishi Heavy Industries (MHI) has develped the "MACT" process and a low-NO_x "PM" burner. "Reburning" or clean fuel overfiring techniques have been developed to destroy NO_x formed in other furnace regions. Reburning has been reported to produce a 50 percent further reduction beyond that of low-NO_x burners. EPRI has sponsored the development of a primary combustion furnace concept with Babcock & Wilcox and is currently sponsoring several retrofit development programs outlined in the next section. EPA has developed the distributed mixing low-NO_x burner that is expected to be demonstrated on a full scale unit in the near future. These continuing developments in the U.S. and Japan hold considerable promise for retrofit NO_x control modifications in the next decade.

2.2.1 Costs, Operational Constraints, and Selection Criteria

Estimates of retrofit capital costs for various control measures by boiler type range from $0.40 to $14.30 per kilowatt as shown in Table 2-1. OFA may require extension of the windbox or ducts to the ports in the furnace wall. Low-NO_x burners may require windbox modifications to accommodate new burner dimensions or to control air flow to the new burners. Consequently, these costs are very site specific with variations of 2 or 3 to 1 not uncommon.

The choice of one combustion modification option over others is strongly dependent upon: (1) capital and operating cost; (2) NO_x reduction potential; (3) physical compatibility of the furnace design with the proposed modification; (4) impact on boiler operational performance (if any); (5) outage requirements (if any); and (6) site-specific factors such as coal type, mill capacity, fan capacity, furnace exit temperature, etc.

One of the primary factors in NO_x emission control is the initial furnace design. The effectiveness of retrofit NO_x control options are directly related to the five major furnace designs, more than five major burner types, unit size, the coal type (bituminous, subbituminous, lignite) and unit age. If one considers the numerous boiler design configurations currently in use, the ten combustion modification options available, and the list of six selection criteria listed above, it should be apparent that the matrix of parameters affecting the choice of an option is very large. It is for this reason that EPRI sponsored a study of NO_x retrofit options that will lead to the preparation of a guideline manual to assist the utilities in making these decisions.

2.3 NEW UNIT VS. RETROFIT NO_x CONTROL CONSIDERATIONS

An important consideration in estimating the potential NO_x reduction that could be achieved by applying NO_x controls on a national basis is the fraction of the units that are eligible for retrofit. This is illustrated in Figure 2-5 where the NSPS units have been segregated because they already have controls and emissions considerably less than their capacity fraction. Tentatively, 60 percent of the total coal-fired NO_x emissions could be considered a candidate for some form of retrofit NO_x controls. However, this does not mean that all units could accommodate retrofit controls because of many other factors that must be considered. For each combination of boiler type, coal type, control method, and site-specific constraint, there is an applicability factor or fraction of eligible units that in concept could be retrofitted. For instance, for the case of commercially available low-NO_x burner, Table 2-2 indicates that only 38 percent of the pre-NSPS coal-fired boiler capacity can be retrofitted.

The importance of physical compatibility with the furnace design was noted previously but deserves further elaboration. This can be illustrated

COAL-FIRED UTILITY BOILERS WITH CAPACITY IN EXCESS OF 100 MW

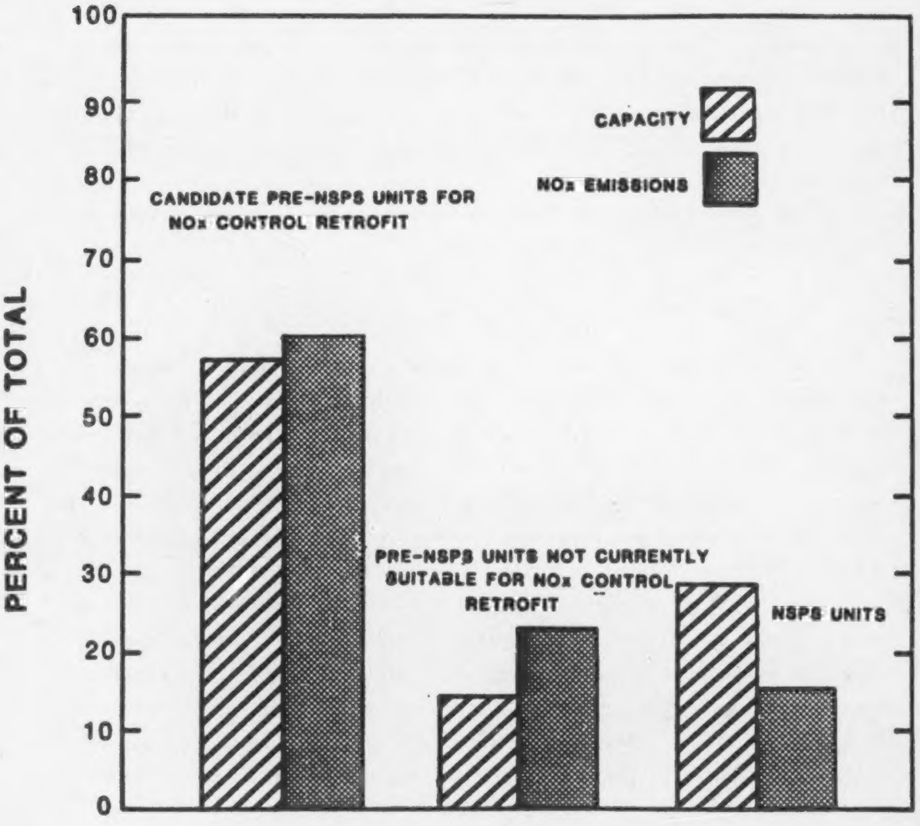


Figure 2-5.

TABLE 2-2.

PRELIMINARY ESTIMATES OF COMMERCIAL LOW-NO_X BURNER RETROFITABILITY FOR COAL-FIRED BOILERS

Boiler Category	Capacity (MW)	% of Total Capacity	Boilers Suitable for Low-NO_x Burner Retrofit %	Boilers Suitable for Low-NO_x Burner Retrofit Capacity
Pre-NSPS Boilers				
CE Tangential	75,958	31	50	37,979
B&W Cell-Burner	24,501	10	-0-	-0-
Cyclone Boilers	23,392	9	-0-	-0-
B&W Wall-Fired w/ Circular Burners	16,280	6	70	11,396
F/W Wall-Fired	15,737	6	70	11,016
R/S Wall-Fired	8,283	3	80	6,626
Wet Bottom Wall-Fired	4,917	2	-0-	-0-
Roof-Fired	3,839	2	-0-	-0-
R/S Turbo-Fired	1,988	1	-0-	-0-
Miscellaneous	1,926	1	-0-	-0-
Pre-NSPS Subtotal ---	176,821	71		67,017
Post-NSPS Boilers				
CE Tangential	30,247	12		
B&W Wall-Fired	21,291	8		
F/W Wall-Fired	12,276	5		
R/S Turbo-Fired	7,138	3		
Post-NSPS Subtotal ---	70,952	29		
TOTAL -	247,773	100		

for the popular option of installing retrofit low-NO_x burners. Furnace volume and width must be considered since a small furnace could lead to flame impingement on the opposite wall and a high heat release leading to tube failures. A short furnace with inadequate length to the convective section inlet will lead to flame impingement at the superheater inlet and poor carbon utilization. If the windbox depth or existing burner throat is inadequate to accommodate the retrofit burner, then substantial modifications to the furnace wall and windbox must be made.

The design philosophy for a retrofit NO_x control method includes many factors besides cost and NO_x reduction potential. Ideally, the retrofit option should be adaptable to a wide variety of unit designs. Because of replacement power generation costs, it should require a minimum of unit downtime for installation. New units can be designed for a specific coal type but a retrofit control method should be adaptable to a wide variety of coal types or there should be provisions for subsequent adjustment in the field after installation. The ease of installation and maintenance is critical because control performance cannot be maintained if the equipment cannot be conveniently serviced.

An important category of consideration in retrofit decision is the current operational history of the unit. Problems with slagging/fouling, inadequate fan capacity, high tube temperatures, poor coal or air distribution to the bur[illegible] and tube wall corrosion, are warning signals that must be considere[illegible]fit. These site-specific factors cannot be addressed in a generic [illegible]tudy of retrofit potential.

[illegible]ltho[illegible]umber of retrofit factors that influence a decision may at first app[illegible]rwhelming, it is important to realize that they all do not car[illegible]hting. A primary objective of EPRI research project RP21[illegible]1 is to [illegible]elop the criteria and a method for evaluating retrofit op[illegible]ns. The boiler manufacturers are responding to the challenge of developing options that can satisfy these criteria and recent burner development programs by Riley and CE, achieving 40-60 percent reductions, provide a basis for optimism.

2.4 EPRI RESEARCH AND DEVELOPMENT PROJECTS

EPRI is currently sponsoring two major projects related to retrofit NO_x combustion controls (RP2154 and RP1836). These projects involve several separate contractors including Babcock & Wilcox (B&W), Combustion Engineering (CE), KVB, Mitsubishi Heavy Industries (MHI), and Riley Stoker. The objectives of these two major contracts are summarized in Table 2-3. The work scope and approach to these objectives is outlined for each of the contractors in the subsections that follow.

2.4.1 Guidelines for Retrofit NO_x Combustion Controls (RP2154-1)

KVB (RP2154-1) will assess the potential for achieving a reduction in coal-fired utility boiler NO_x emissions on a national and local basis by applying retrofit NO_x controls. It is a generic assessment study that covers all boilers greater than 100 MW in capacity (which includes more than 90 percent of the annual NO_x emissions).

One of the initial tasks of the KVB retrofit study was the categorization of boilers by design features that influence NO_x emissions. Figure 2-6 is an example of this categorization by firing configuration and manufacturer for three time periods representing significantly different design philosophy. Note that the cyclone and cell-burnered units having an opposed firing configuration represent a substantial fraction the NO_x emissions.

The study will identify potential NO_x retrofit options and evaluate critical design, performance, and operational factors. Estimates are being prepared for NO_x control reduction and costs which will become some of the major factors in control method selection. Many factors enter into the selection of a NO_x control method as shown in Figure 2-7. KVB is developing a methodology for matching the control options best suited to the individual boiler design categories.

Although the KVB study cannot address all site-specific considerations in its projection of retrofit controls application, it will outline some of these considerations in a retrofit guideline document to assist utilities in

TABLE 2-3. RETROFIT LOW-NO$_x$ COMBUSTION CONTROLS FOR COAL-FIRED BOILERS (RP2154)

(Contractors: EVB, Riley-Stoker, and B&W)

Objectives

- Determine the potential for applying retrofit NO$_x$ controls
- Develop a guideline to assist utilities in retrofit decisions
- Develop a design basis for retrofitting low-NO$_x$ burners and advanced fuel/air staging on wall-fired units
- Develop a design basis for retrofitting [illegible]$_x$ burners and advanced staging on cell burner units
- Evaluate commercial feasibility

EVALUATION OF LOW-NO$_x$ BURNER "PM" FOR TANGENTIALLY FIRED UNITS (RP1836-1)

(Contractor: CE and MHI)

Objectives

- Evaluate the NO$_x$ emissions of PM burner firing U.S. coals
- Evaluate the retrofit potential and new unit costs associated with meeting specific NO$_x$ guarantees

DISTRIBUTION OF ANNUAL NOx EMISSIONS BY DESIGN PERIOD, FIRING ARRANGEMENT AND MANUFACTURER

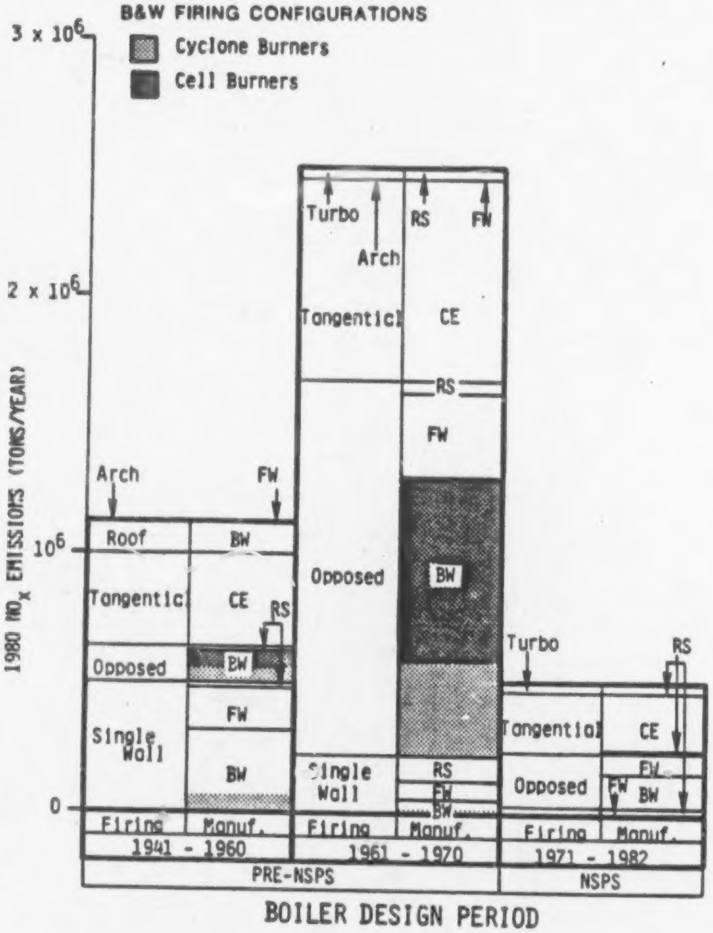
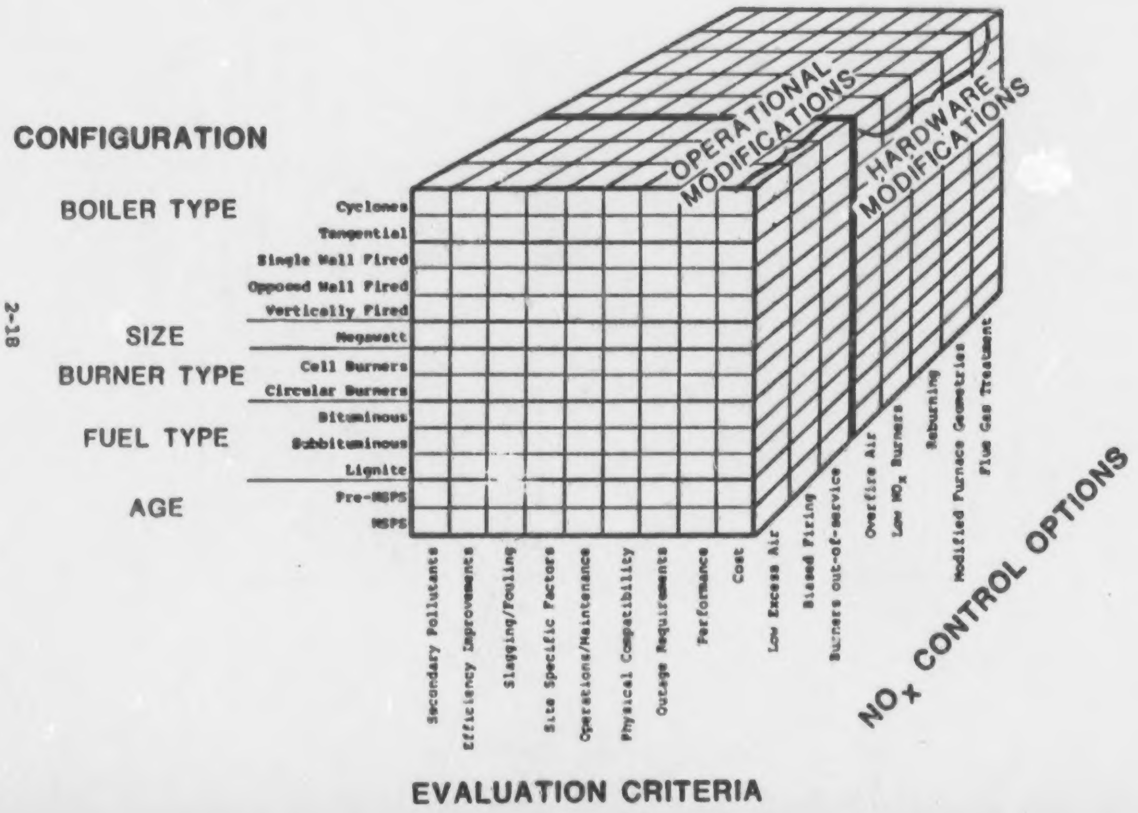


Figure 2-6.

MULTIPLE TRADE-OFFS IN RETROFIT NO_x CONTROLS


CONFIGURATION
BOILER TYPE
Cyclones
Tangential
Single Wall Fired
Opposed Wall Fired
Vertically Fired
SIZE
Megawatt
BURNER TYPE
Cell Burners
Circular Burners
FUEL TYPE
Bituminous
Subbituminous
Lignite
AGE
Pre-NSPS
NSPS
OPERATIONAL MODIFICATIONS
HARDWARE MODIFICATIONS
Secondary Pollutants
Efficiency Improvements
Slagging/Fouling
Site Specific Factors
Operations/Maintenance
Physical Compatibility
Outage Requirements
Performance
Cost
Low Excess Air
Biased Firing
Burners out-of-service
Overfire Air
Low NOx Burners
Reburning
Modified Furnace Geometries
Flue Gas Treatment
NOx CONTROL OPTIONS
EVALUATION CRITERIA

their decisions. Since cost will be a primary concern, cost effectiveness curves will be generated to determine the incremental costs associated with added NO_x control.

2.4.2 Evaluation of Low-NO_x "PM" Burner for Tangentially Fired Units (RP1836-1)

Combustion Engineering is conducting a NO_x engineering study relevant to the tangentially fired units. Mitsubishi Heavy Industries (MHI) has developed a low-NO_x "PM" burner which is licensed to Combustion Engineering for application to U.S. boilers. One of the primary objectives of this study is to evaluate the NO_x emissions of the PM burner on U.S. coals and the retrofit costs relative to meeting specific NO_x guarantees.

The PM burner is a unique firing system (Figure 2-8) consisting of alternate fuel-rich and fuel-lean zones that minimize the conversion of fuel bound nitrogen to NO_x. The current CE study evaluates four options of overfire air compartment design for the single and divided furnace tangentially fired boilers. NO_x reductions between 45 and 60 percent have been projected for full-scale retrofits as shown in Table 2-4. Retrofit costs ranging from 8 to 16/kW are dependent upon the furnace configuration and the complexity of the retrofit.

Because of the interest in combined NO_x/SO_2 control, MHI and CE are also performing limestone injection tests to assess the SO_2 removal potential of the PM burner system. These tests are being conducted with both Japanese and American coals to establish correlations to prior test data on Japanese coals. Subscale low-NO_x burner/limestone injection tests will study the dependence of emissions on coal properties, injection rate and method of injection. Tangential units have had attractive NO_x emissions which potentially can be reduced further with the PM burner. If a successful system can be developed integrating limestone injection for SO_2 removal, the tangential units will have the potential for substantial reductions in NO_x and SO_2 emissions. Combined NO_x and SO_2 removal processes are discussed in Section 3.0.

PM CELL IN TANGENTIAL FIRING SYSTEM

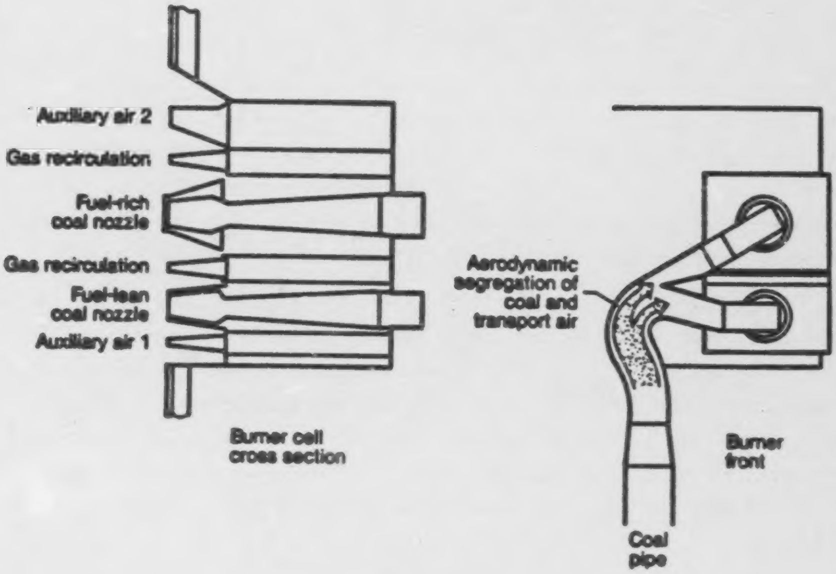

Auxiliary air 2
Gas recirculation
Fuel-rich coal nozzle
Gas recirculation
Fuel-lean coal nozzle
Auxiliary air 1
Burner cell cross section
Aerodynamic segregation of coal and transport air
Burner front
Coal pipe

Figure 2-8.

TABLE 2-4.

PM BURNER RETROFIT

NOx PERFORMANCE AND ECONOMIC SUMMARY[3]

	NO_x GUARANTEE[1]	% NO_x REDUCTION[2]	TOTAL CAPITAL $/KW	POWER COSTS MILLS/KW-HR
SINGLE FURNACE				
MINIMAL COST CASE	190	46	7.91	0.22
HIGHEST COST CASE	150	57	15.90	0.45
DIVIDED FURNACE	160	50	14.29	0.40
NEW UNIT	150	57	5.44	0.15

[1] PPM, CORRECTED TO 3% O_2

[2] BASED ON SINGLE FURNACE EMISSIONS OF 350 PPM
BASED ON DIVIDED FURNACE EMISSIONS OF 320 PPM

[3] 1981 DOLLARS CONFIDENCE LEVELS 20%

2.4.3 Evaluation of Retrofit NO_x Combustion Controls for Wall-Fired Utility Boilers (RP2154-2)

Riley-Stoker (RP2154-2) is developing retrofit low-NO_x burners and advanced fuel/air staging for wall-fired units. Riley-Stoker has developed controlled combustion venturi (CCV) burner that has shown considerable prom for wall-fired retrofit applications. The main features of this burner (Figure 2-9) are the venturi nozzle tip and four-bladed conical coal spread at the exit of the primary air/coal pipe. This arrangement establishes loc fuel-rich controlled combustion zones that have produced NO_x reductions of to 60 percent as shown in Figure 2-10.

The Riley-Stoker work will continue with an evaluation of advanced control systems using the CCV burner in combination with furnace staged combustion. A primary objective associated with the process development will the identification and resolution of major commercialization issues. Pilot-scale testing will be conducted to confirm the performance potential followe by economic assessment and application studies. The current program will conclude with a planning and test strategy task for a possible full-scale demonstration in a subsequent program.

2.4.4 Evaluation of Retrofit NO_x Combustion Controls for Cell-Burner Units (RP2154-3)

B&W is conducting a similar retrofit low-NO_x burner and advanced staging development effort but specifically for cell burner units. As note previously, cell burner units represent a substantial fraction of total util ity NO_x emissions and no retrofit burner options are commercially available this time. After evaluating potential retrofit options for these units, B& will conduct 6 MBtu pilot-scale burner development testing and associated h transfer and process scale-up engineering evaluations. Important issues concerning slagging, fouling, and corrosion will be addressed followed by 1 MBtu/hr scale testing and final application studies.

Although this program possibly poses a more difficult challenge than the prior contracts, it is essential to providing a retrofit NO_x control option to an important category of units for which no option is presently available.

CONTROLLED COMBUSTION VENTURI (CCV) BURNER

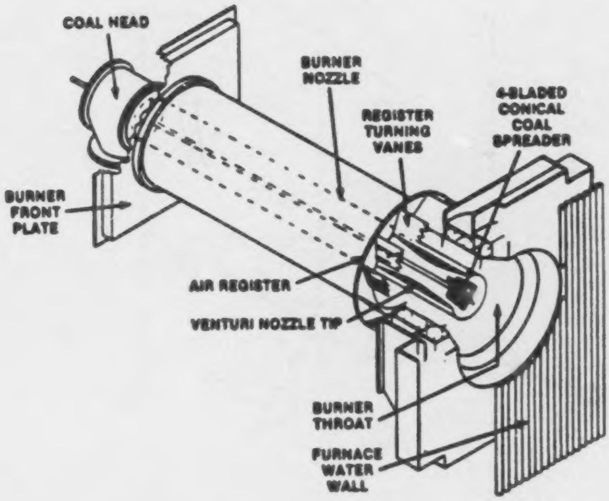

COAL HEAD
BURNER NOZZLE
REGISTER TURNING VANES
4-BLADED CONICAL COAL SPREADER
BURNER FRONT PLATE
AIR REGISTER
VENTURI NOZZLE TIP
BURNER THROAT
FURNACE WATER WALL

Figure 2-9.

NOx REDUCTION WITH RETROFIT CCV BURNER

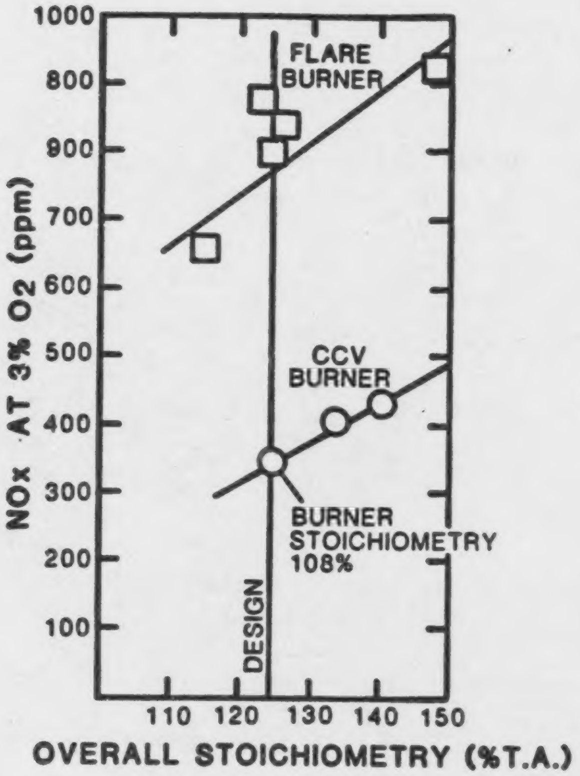


Figure 2-10.

SECTION 3.0

DRY SORBENT INJECTION FOR SO_2 CONTROL

Dry processes, both in-furnace (calcium-based sorbent) and post-furnace (sodium-based sorbent), for SO_2 control were discussed in Section 1.0. While neither process is a new technology, each having undergone development in the 1960's, the criteria under which they are being evaluated and applications considered have changed over the ensuing years. To assess the current state and future prospects of dry sorbent injection for SO_2 control requires not only a better understanding of process fundamentals and an assessment of the potential SO_2 removals, but also potential impacts on boiler performance must be determined. EPRI has a comprehensive program addressing these issues for both in-furnace and post-furnace sorbent injection for SO_2 control.

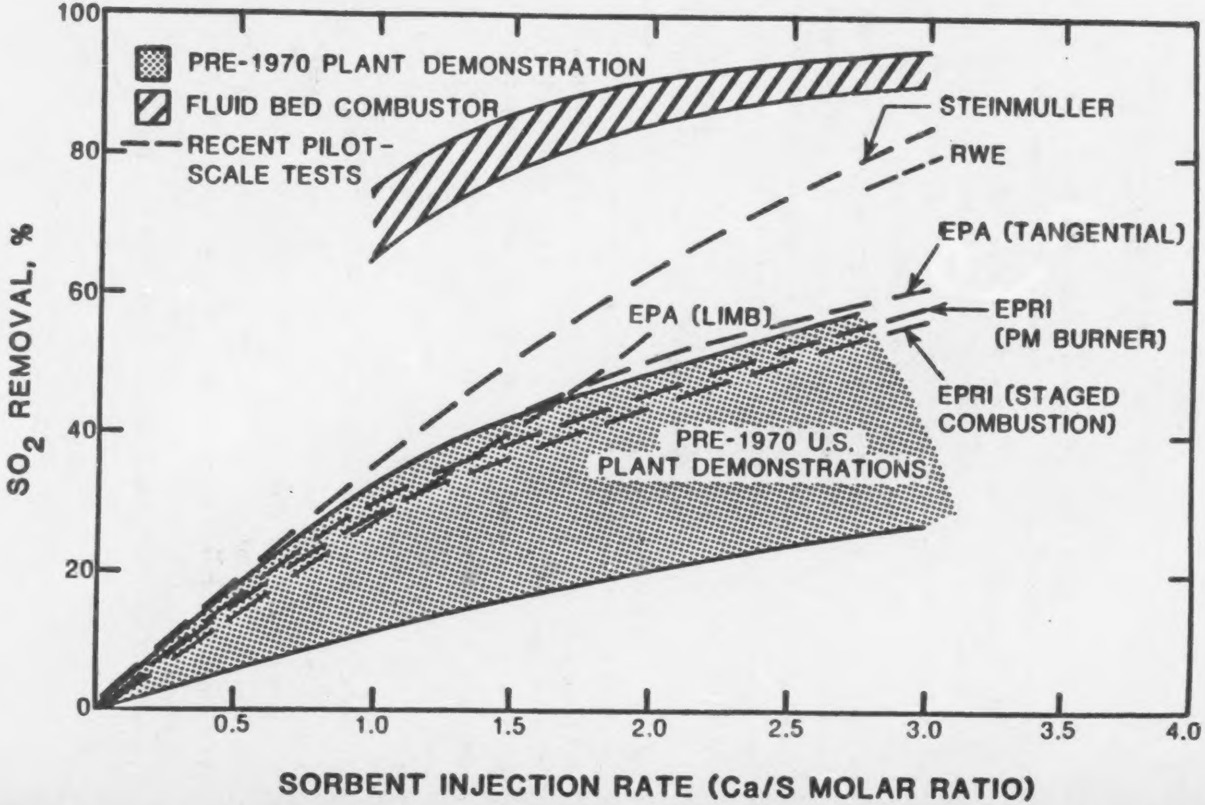
In-furnace sorbent injection presently appears to be more widely applicable than post-furnace sorbent injection primarily due to the availability and cost of the dry sorbents. Thus, this section of the presentation will emphasize furnace sorbent injection technology, and only provide an overview of post-furnace sorbent injection. However, it is recognized that development of new sorbent materials for post-furnace injection could potentially expand the applicability of these processes in the future.

3.1 FURNACE SORBENT INJECTION

3.1.1 Historical Perspective on Furnace Sorbent Injection

In-furnace dry sorbent injection for SO_2 control was investigated during the 1960's and early 1970's in the United States, Europe, and Japan and included and a number of full-scale demonstrations. These full-scale demonstrations were conducted in a variety of boiler designs and for practical sorbent injection rates yielded SO_2 removals in the range of about 20-50 percent. This range of SO_2 removals is shown as the shaded area in Figure 3-1

HISTORICAL PERSPECTIVE ON IN-FURNACE SORBENT INJECTION FOR SO_2 CONTROL



3-2

as a function of sorbent injection rate. Sorbent injection rate is generally expressed as a fraction of the amount needed to theoretically react with all of the SO_2 in the flue gas (e.g. Ca/S molar ratio of unity). The SO_2 removals that were achieved were dependent not only on the sorbent type and injection rate but also on the boiler design. Higher SO_2 removals were associated with boiler designs that were characterized by gradual fuel/air mixing and distributed heat release rates (e.g., arch-fired boilers with sidewall air addition and tangentially fired boilers). In fact, the upper portion of the band of SO_2 removals shown in Figure 3-1 is defined by the results from an arch-fired boiler. It should be noted that arch-fired boilers, which by design, exhibit gradual mixing of the coal and air have been documented in a recent EPRI Project RP-1339-1 to exhibit low-NO_x emission characteristics. As another point of reference, SO_2 removals with limestone addition to fluid bed combustors is also shown in Figure 3-1 since the overall chemistry is comparable to in-furnace sorbent injection.

The relatively low SO_2 removals and the variability from boiler to boiler during this early development effort is attributed to three factors:

- "dead burning" of the CaO, when the sorbent was injected into the burner/flame zone which decreased the reactivity of the sorbent
- inadequate mixing of the sorbent wth the combustion products when the sorbent was injected away from the flame zone in the upper furnace region
- poor fundamental understanding of the SO_2 capture process precluding process optimization for a site-specific boiler burner system

Because the SO_2 removals were well below the 80-90 percent target values and variable, the process was not considered a technical success and further process development work was abandoned in the early 1970's. At this point, the development of wet scrubber systems became the primary focus of SO_2 control efforts based on the potential SO_2 removals and overly optimistic cost estimates of the wet scrubbing systems.

There has been a renewed interest in furnace sorbent injection as a result of recent pilot-scale test results from the United States, Europe, and

Japan as well as results from a 60 MW brown coal-fired boiler. These recent results indicate the potential to achieve higher SO_2 removal efficiencies at practical sorbent injection rates, provided combustion and sorbent conditions can be properly controlled. Also, these latest results indicate that the thermal environment associated with combustion systems being developed for low-NO_x emissions contribute to optimizing SO_2 capture. In particular, the lower combustion temperatures and controlled fuel air mixing which is characteristic of the majority of the low-NO_x combustion systems may provide a new opportunity to optimize the sorbent injection process. SO_2 removals of 50 percent may be possible at reasonable sorbent injection rates. Results from these recent pilot-scale programs are also included in Figure 3-1.

3.1.2 Process Description

Furnace sorbent injection for SO_2 control consists of injecting a dry pulverized calcium-based material directly into the furnace cavity of a pulverized coal-fired boiler. The process is illustrated in Figure 3-2. Upon entering the furnace, the material converts to lime (CaO) which reacts in suspension with SO_2 to form solid calcium sulfate. The calcium sulfate and unreacted sorbent material is then removed along with the fly ash in the conventional particulate control device (electrostatic precipitator or fabric filter).

The process is entirely dry and utilizes conventional equipment for storage, handling, and injection of the sorbent material. The sorbent materials of interest are calcium-based compounds including limestone ($CaCO_3$), dolomite ($CaCO_3MgCO_3$), calcium hydroxide ($Ca(OH)_2$), and possibly lime (CaO). Limestone is the most abundant material available throughout the United States and is the preferred dry sorbent material from the standpoint of availability and cost.

The furnace sorbent injection process requires that the SO_2 removal process steps be integrated within a conventional furnace system, as opposed to adding a separate pollution control device to the utility boiler for SO_2 removal. The lack of success in the early development of this technology can be attributed to an insufficient understanding of the process to effectively implement this integration on a site-specific basis. For instance, the recent

DRY SORBENT FURNACE INJECTION PROCESS

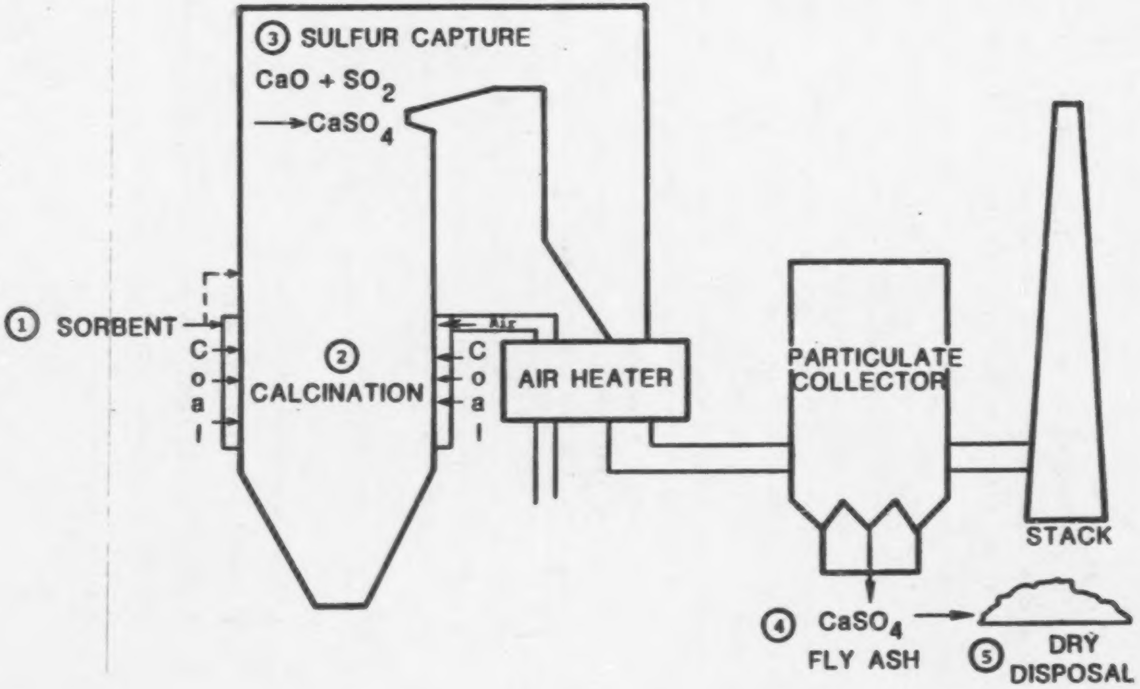

③ SULFUR CAPTURE
CaO + SO2
→CaSO4
① SORBENT
C
o
a
l
② CALCINATION
C
o
a
l
Air
AIR HEATER
PARTICULATE COLLECTOR
STACK
④ CaSO4
FLY ASH
⑤ DRY
DISPOSAL

Figure 3-2.

pilot-scale results point to the requirement that the thermal environment in the near-burner region be carefully controlled to increase SO_2 removal. Currently, low-NO_x burners appear to be the only reasonable near-term approach to control mixing and combustion in the near-burner region of the furnace in order to optimize SO_2 removal by in-furnace sorbent injection (see Table 2-2).

The process involves a heterogeneous gas-solid reaction between the sorbent particle and SO_2. This reaction process, as illustrated in Figure 3-3 for limestone, depends not only on the chemical nature of the sorbent particle, but is also dependent on the physical properties of the injected sorbent, and the evolution of the physical properties in the furnace. Upon entering the furnace the sorbent material heats up and calcines resulting in the evolution of CO_2. The evolution of CO_2 creates a high surface area, porous structure that enhances SO_2 capture. As reaction with SO_2 proceeds, a layer of calcium sulfate builds up which will ultimately coat the surface preventing further reaction with SO_2. The extent to which an open pore structure can be maintained, resulting in high SO_2 capture and high sorbent utilization, depends not only on the properties of the injected sorbent but also on interaction between the sorbent and the thermal environment in the furnace. To confidently apply the process, particularly to retrofit applications, it is necessary that a sufficient understanding of the process be developed to understand and accommodate the key process parameters which include:

- Furnace Design (Temperature/Time History, Sorbent Injection Location)
- Coal Sulfur Content
- Sorbent Type (Limestone, Dolomite, Calcium Hydroxide)
- Sorbent Physical Properties (Surface Area, Pore Structure)
- Sorbent Injection Rate (Calcium/Sulfur Ratio)
- Sorbent Particle Size

FURNACE SORBENT INJECTION PROCESS FUNDAMENTALS

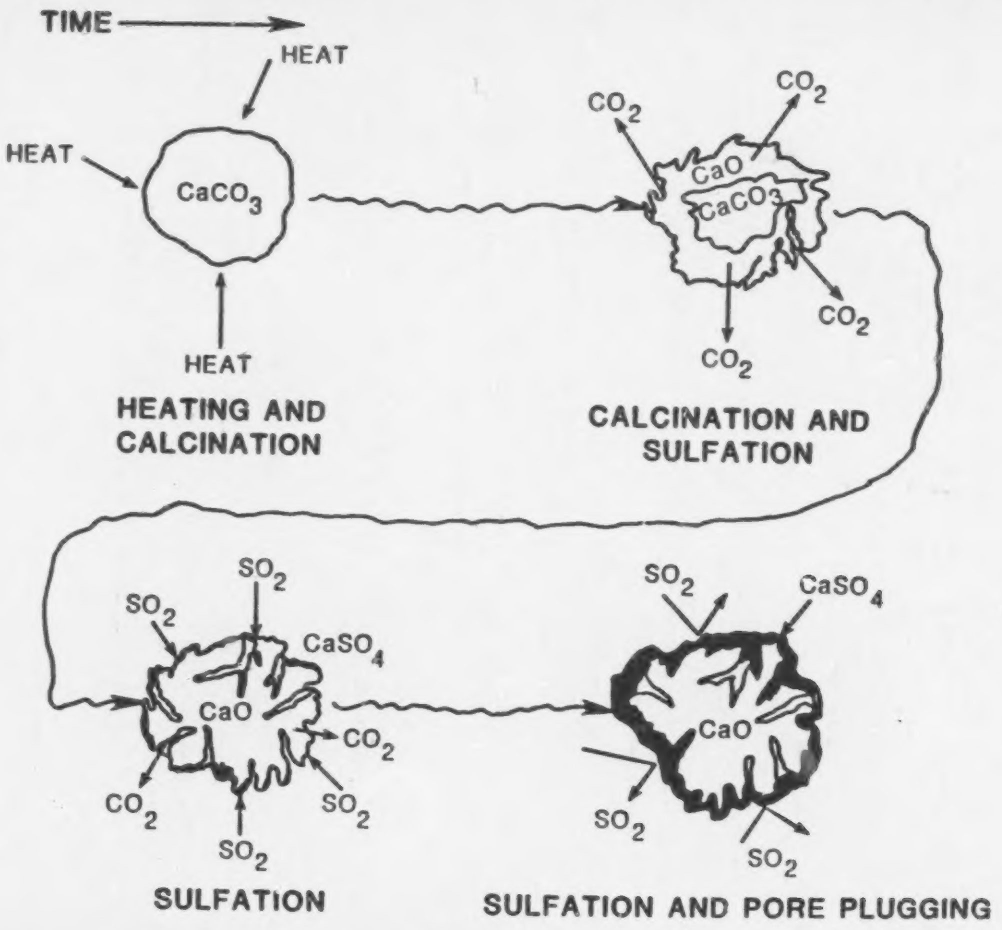

TIME
HEAT
HEAT
$CaCO_3$
HEAT
HEATING AND CALCINATION
CO_2
CO_2
CaO
$CaCO_3$
CO_2
CO_2
CALCINATION AND SULFATION
SO_2
SO_2
$CaSO_4$
CaO
CO_2
CO_2
SO_2
SO_2
SULFATION
SO_2
$CaSO_4$
CaO
SO_2
SO_2
SULFATION AND PORE PLUGGING

Figure 3-3

3.1.3 Advantages and Applicability

The potential SO_2 removals with furnace sorbent injection are not sufficient to meet current New Source Performance Standards on high sulfur coals. However, there are a number of retrofit applications where incremental SO_2 control via furnace sorbent injection will be logical to minimize overall SO_2 control cost. These include applications to:

- Supplement coal cleaning
- Augment existing flue gas desulfurization systems
- Allow purchase of cheaper, high sulfur coal in meeting State Implementation Plans, etc.

The potential advantages of furnace sorbent injection in such applications can be summarized as follows:

1. Retrofittability: The sorbent injection process can be retrofitted to existing coal-fired boilers using commercially available equipment.
2. Simplicity: The relative simplicity of the process compared to conventional wet chemical processes minimizes the power plant operation and maintenance burdens, and should lead to higher boiler availability.
3. Cost: Low capital and operating costs compared to wet flue gas treatment. (More than $100/kW less in capital requirement and levelized busbar costs potentially one-half that of wet scrubbing.)

3.1.4 Developmental Status and Research Activities

While the early development work did not provide an optimistic outlook for the technology, recent pilot-scale work indicates that further assessment and development of the technology is warranted. Major research efforts to develop furnace sorbent injection are ongoing in the United States, Europe, and Japan.

A. International Development Programs--

The most extensive international development program is being conducted by West German utilities and boiler manufacturers. Steinmuller, a major German boiler manufacturer, is developing furnace sorbent injection for wall-fired boilers using a unique low-NO_x burner design and coals similar to eastern, low sulfur (1 percent), bituminous coals. The burner, which controls mixing and combustion in the furnace flame zone, is reportedly essential to achieving high SO_2 removals with low sulfur bituminous coals. This work entailed an extensive 2.3 MW (thermal input) pilot-scale effort at the International Flame Research Foundation with demonstrated SO_2 removals of 80 percent at sorbent injection rates corresponding to a Ca/S molar ratio of 3 to 1 (Figure 3-1). Steinmuller has chosen to proceed from the pilot-scale to a 700 MW full-scale demonstration at the Weiher Number 3 unit of Saarbergwerke near Saarbrucken, West Germany to address full-scale SO_2 removal. This full-scale work is primarily aimed at assessing sorbent solids handling and assessing potential boiler impacts, as opposed to optimizing SO_2 removal. This full-scale demonstration is currently in the start-up phases and no results are available. The potential applicability of the Steinmuller approach in this country may hinge on the retrofittability of the burner to U.S. utility boiler designs.

The other major development program in West Germany is being performed by the electric utility company Rheinisch-Westfalisches Electrizitatswerk (RWE). RWE's work has involved both laboratory and small utility programs with demonstrated SO_2 removals over 60 percent on a 60 MW brown coal-fired boiler (Figure 3-1). Brown coal-fired boilers are characterized by low furnace peak gas temperatures (less than 2100°F) due to the 40-60 percent moisture content of the fuel. Due to the unique properties of the coal and the particular design of the boilers (large furnace volumes and residence times), the combustion conditions appear to be especially well suited for in-furnace SO_2 capture using sorbent injection. Although, since these conditions are not typical of U.S. boilers, these results may not be directly applicable to utility boiler systems in the United States. However, the results do provide a benchmark for the potential of furnace sorbent injection. RWE is currently scaling the process to a 300 MW unit.

Ongoing work in Japan, sponsored by EPRI (RP 1836-1) involves an exploratory investigation of incorporating furnace sorbent injection along with the low-NO_x PM burner (see Section 2.0 for description of PM burner).

Preliminary results from the study indicate the potential for SO_2 removals of 50 percent at Ca/S molar ratios of 2 to 1.

Development programs of furnace sorbent injection are also being pursued in other countries including Austria and Canada.

B. United States Development Activities--

In the United States development is principally supported by EPRI and EPA. To date research has focused on laboratory programs to develop process data and to evaluate potential power plant impacts prior to committing to full-scale demonstrations. EPA projects have included laboratory experimental programs and technical and economic analyses aimed at establishing designs and costs for sorbent injection systems for wall-fired and tangentially fired boilers. For wall-fired boiler applications, the sorbent injection process is coupled with a low-NO_x burner similar to the Steinmuller burner design mentioned previously. The combined NO_x/SO_2 control system, referred to as "LIMB" (limestone injection multistaged burner), is intended for retrofit to coal-fired boilers. Reported development targets for this process are 50 to 70 percent SO_2 control. Results of pilot-scale studies sponsored by EPA are also included in Figure 3-1.

EPRI has an ongoing program aimed at developing furnace sorbent injection process data as well as resolving boiler impact issues and establishing the relative economics of the process. The EPRI projects include:

RP 1836-3	A preliminary engineering and economic feasibility study of retrofit furnace limestone injection
RP 1836-1	Pilot-scale proof-of-concept testing involving limestone injection in combination with a low-NO_x burner
RP 899-2	Laboratory fouling and slagging tests to measure potential boiler fireside impacts
RP 899-2-2	Exploratory laboratory tests to evaluate the possible reuse of unreacted furnace sorbent material for supplemental postcombustion SO_2 removal
RP 1682-1	Economic comparison of furnace limestone injection and dry flue gas treatment for SO_2 controls

These projects have set the groundwork for a major expansion of the EPRI program which will include utility companies' participation in development testing, prototype design, and field demonstrations. This development schedule is discussed in more detail in Section 3.1.6.

3.1.5 Developmental Issues

The need to develop a better knowledge of the process has already been emphasized. Additionally, there is a need to establish the impacts that furnace sorbent injection will have on the performance of the boiler system. These potential impacts are illustrated in Figure 3-4 and include:

- convective section erosion
- increased particulate loadings
- slagging
- convective section erosion
- retrofittability of a low-NO_x system
- air heater pluggage
- particulate collector performance
- ash handling
- furnace corrosion

Commercial application of this technology cannot proceed on a utility-wide basis until the above impacts are assessed and resolved. Resolution of these issues has been a major part of EPRI's program in this technology area. The following paragraphs will briefly discuss the status of each of these issues.

Particulate Loading

Furnace sorbent injection will impact the total particulate loading that must be handled by the utility boiler system. This increased loading can potentially impact slagging, fouling, erosion, air heater deposition, particulate collection and ash handling. Figure 3-5 illustrates this impact for a typical 10 percent ash coal for sorbent (limestone) injection rates corresponding to calcium to sulfur ratios of 1 to 4. As can be seen in Figure 3-5, the impact on solids loading is both highly dependent on coal sulfur content as well as the Ca/S ratio. For instance, a Ca/S requirement of 2 to 1 for SO_2 removal from a boiler firing a coal with a sulfur content of two percent would double the solid loadings in the system. The ability of the system to handle these increased loadings must be addressed.

POTENTIAL RETROFIT IMPACTS OF FURNACE SORBENT INJECTION

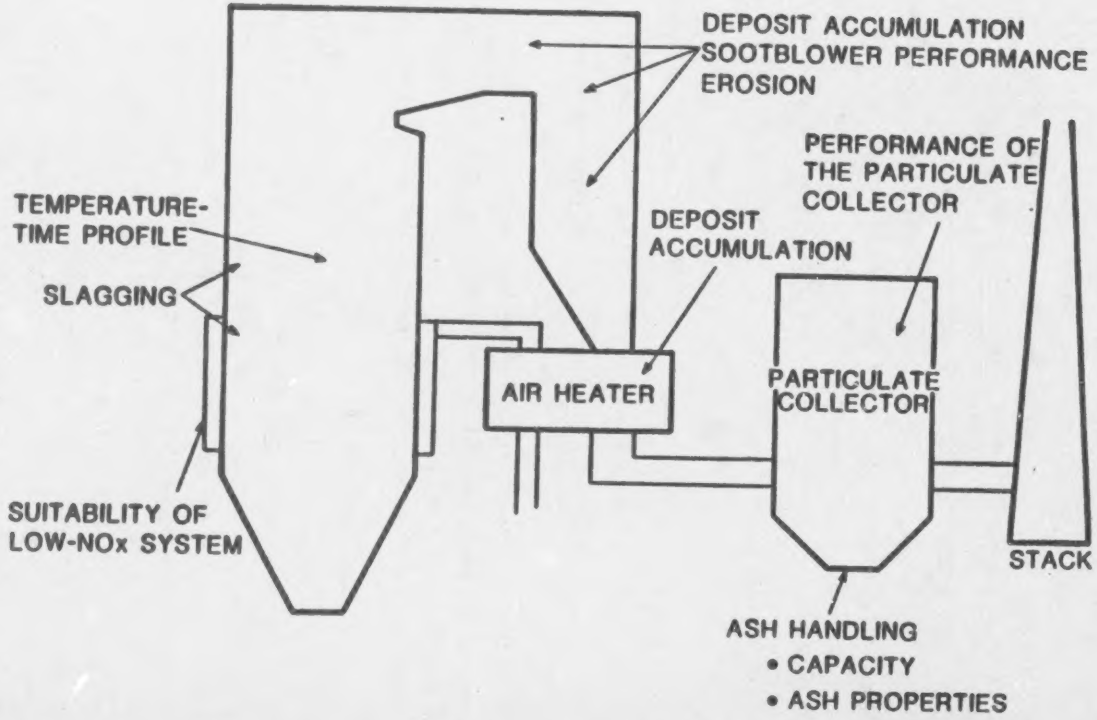

DEPOSIT ACCUMULATION
SOOTBLOWER PERFORMANCE
EROSION
PERFORMANCE OF THE PARTICULATE COLLECTOR
TEMPERATURE-TIME PROFILE
DEPOSIT ACCUMULATION
SLAGGING
AIR HEATER
PARTICULATE COLLECTOR
SUITABILITY OF LOW-NOx SYSTEM
STACK
ASH HANDLING
• CAPACITY
• ASH PROPERTIES

Figure 3-4.

INCREASED BOILER SOLIDS LOADING FOR FURNACE SORBENT INJECTION

(TYPICAL 10% ASH COAL)

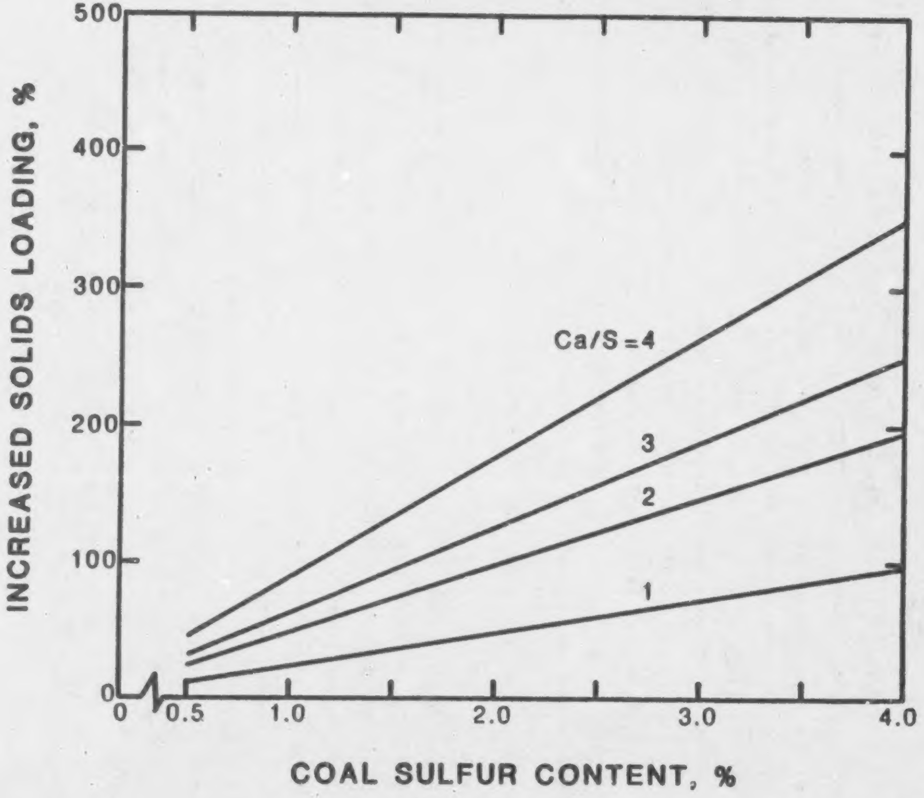


Figure 3-5.

Furnace Slagging

Furnace sorbent injection for SO_2 control will not only impact the particulate loading in the boiler but can also alter the chemical composition of the ash. The change in chemical composition can potentially impact slagging in the radiant section of the furnace. The impact of sorbent injection on slagging is the focus of EPRI Project RP-899-2 being performed in Combustion Engineering's Fireside Performance Test Furnace. The preliminary analysis of the data from this program indicates a potential favorable impact on slagging. While the thickness of the slag layer was found to increase with furnace sorbent injection, the slag deposit that accumulated was easier to remove and manageable with conventional equipment.

Convective Fouling and Deposition

The impact of furnace sorbent injection on convective section particulate deposition and fouling is another issue that has been the subject of EPRI Project RP-899-2. The results have shown that with both eastern bituminous and western bituminous coals the rate of deposition on a simulated tube section increases. This is illustrated in Figure 3-6 showing the sootblowing interval required to maintain a deposit buildup of less than a specified thickness. Without sorbent injection, sootblowing was required every 16-20 hours. The sootblowing frequency had to be increased to approximately once every eight hours at a sorbent injection rate corresponding to Ca/S = 4. While the sootblowing frequency had to be increased, the deposits were easily removed. In fact as shown in Figure 3-7 for the case of a western subbituminous coal, sorbent injection resulted in a substantial reduction in the deposit bond strength from a condition where it was difficult to remove the deposits to a condition where sootblowing was effective.

The economizer is another area that can be impacted by furnace sorbent injection. Application of this technology to units that have finned tube economizers will likely require modifications to the economizer to prevent excessive deposit accumulation.

IMPACT OF FURNACE SORBENT INJECTION ON CONVECTIVE SECTION SOOTBLOWING REQUIREMENT
(PILOT-SCALE RESULTS)

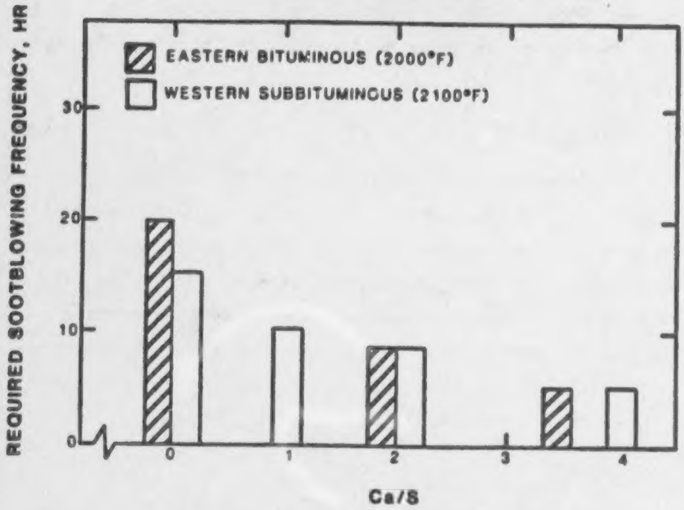


Figure 3-6.

IMPACT OF FURNACE SORBENT INJECTION ON CONVECTIVE SECTION DEPOSIT STRENGTH
(PILOT-SCALE RESULTS)

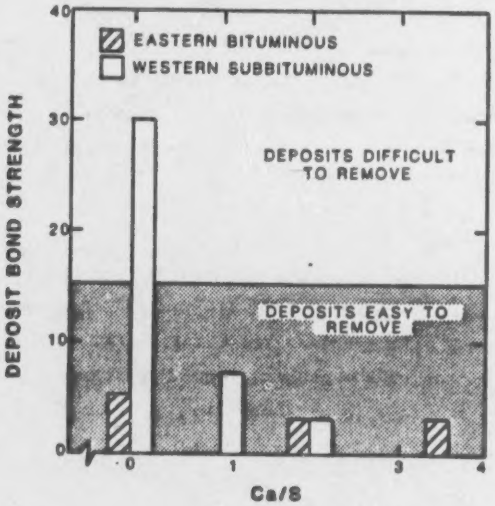


Figure 3-7.

Furnace Corrosion

No definitive data is currently available to assess the potential impact of furnace sorbent injection on corrosion. This assessment is complicated by the extensive test times required to obtain statistically meaningful corrosion data.

Low-NO_x Combustion System

The recent pilot-scale results (RP-1836-1, RP-899-2) indicate the possibility of increasing SO_2 capture through the use of a low-NO_x combustion system. This may be necessary to achieve potential SO_2 removal levels with high sulfur bituminous coals. If this is substantiated by further process development work, then the feasibility of retrofitting a furnace sorbent injection system to a utility boiler will be dependent on the suitability of incorporating a low-NO_x combustion system into the boiler. These retrofit issues of low-NO_x systems were discussed in Section 2.0.

Air Heater Deposition

Currently, there is no definitive data on the potential for particulate deposition and pluggage due to sorbent injection. Further work is warranted to assess this potential impact.

Performance of the Particulate Collector

The performance of an existing electrostatic precipitator subject to a furnace dry sorbent injection retrofit needs to be addressed. The collection efficiency due to the higher solids loadings and potentially higher resistivity ash needs to be resolved. Currently, it is uncertain whether (1) the performance will be affected; (2) a modification can be made to upgrade the performance; or (3) the system requires a fabric filter retrofit or another electrostatic precipitator.

Ash Handling Issues

The capacity and suitability of the ash removal and handling systems need to be addressed on a site-specific basis for retrofit application. The high calcium sulfate content of the ash may result in cementitious properties that can cause plugging problems in systems using wet sluicing ash removal.

3.1.6 Development Schedule

The process development and commercialization schedule for high sulfur U.S. coal is shown in Figure 3-8. This schedule assumes a step-wise progression involving: (1) laboratory development ending in late 1984, (2) prototype verification testing of a small (20-80 MW) utility boiler in 1986, and (3) the first large (400+MW) utility boiler demonstration beginning in 1988. This step-wise approach will enable confident scale-up of the process from each stage in the development, thus providing a maximum probability of success and least risk to host boiler equipment. Interim results may permit application of the process earlier to some existing power plants where process optimization is not required.

The Steinmuller 700 MW field testing with low sulfur bituminous coal is scheduled to begin in late Summer 1983. Another field demonstration in Germany with "brown" coal by RWE at 300 MW scale was also begun in January 1983. Assuming that the plant system impact issues can be successfully resolved at full scale, it is possible that these programs can accelerate the development and demonstration efforts in the U.S. However, due to the unique boiler design, coals, low sulfur content, burners and sorbent materials involved in the West German development programs, the direct application of results to the U.S. utility industry may be limited.

3.1.7 Process Economics

Because of the developmental status of the technology, no actual process costs are available. However, retrofit cost estimates have been made for the furnace sorbent injection process using EPRI's standard economic and design premises. As shown in Figure 3-9, retrofit capital requirements range from $14 to $50/kW depending on the sulfur content of the coal and the required sorbent injection rate. These capital costs include (1) the sorbent storage, handling and injection system, and (2) provision for a low-NO_x combustion system which may vary in cost from 1-15 $/kW depending on boiler and burner design, and (3) economizer modifications to accommodate the higher particulate loadings ($3.8/kW for a 500 MW unit). The capital costs do not include any modifications to the particulate collection system or ash handling

FURNACE SORBENT INJECTION DEVELOPMENT AND COMMERCIALIZATION SCHEDULE

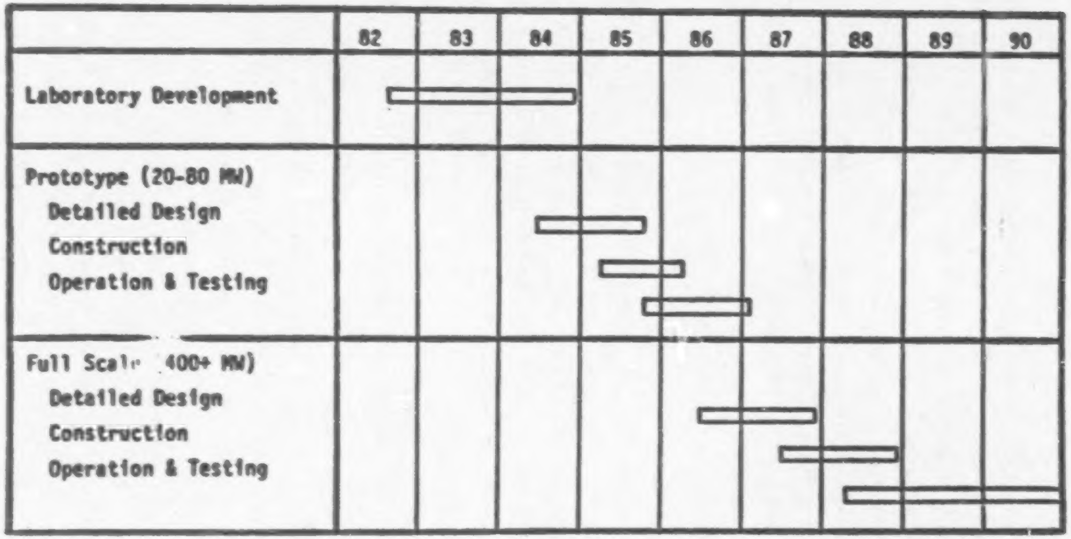


Figure 3-8.

ESTIMATED CAPITAL COSTS OF FURNACE SORBENT (LIMESTONE) INJECTION

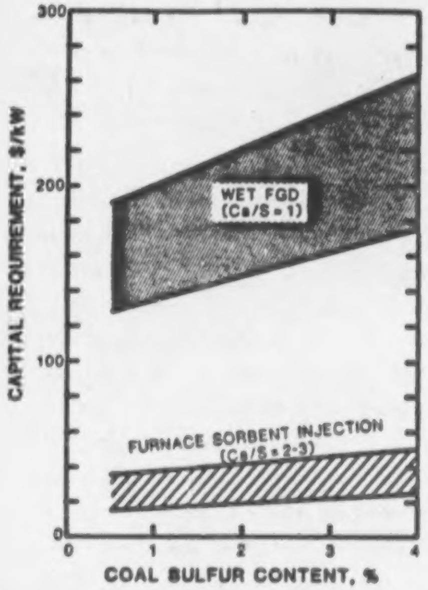


Figure 3-9.

COST ASSUMPTION

WET FGD

- CAPABILITY FOR 90% SO_2 REMOVAL
- CONVENTIONAL WET LIMESTONE SCRUBBER
- RETROFIT CAPITAL COST 25% HIGHER THAN NEW UNIT APPLICATIONS
- PLAN AREA AVAILABILITY
- CONVENTIONAL WET O&M COSTS

FURNACE SORBENT INJECTION

- CAPABILITY FOR 50% SO_2 REMOVAL
- LIMESTONE SORBENT
- UP TO 15 $/kW FOR LOW NOx BURNER RETROFIT
- PARTICULATE CONTROL UPGRADING AND SPECIAL PREDISPOSAL TREATMENT OF BYPRODUCT NOT REQUIRED
- SLAGGING/FOULING MANAGEABLE WITH SOOTBLOWER REDESIGN AND ECONOMIZER MODIFICATION

ESTIMATED LEVELIZED BUSBAR COSTS OF FURNACE SORBENT (LIMESTONE) INJECTION

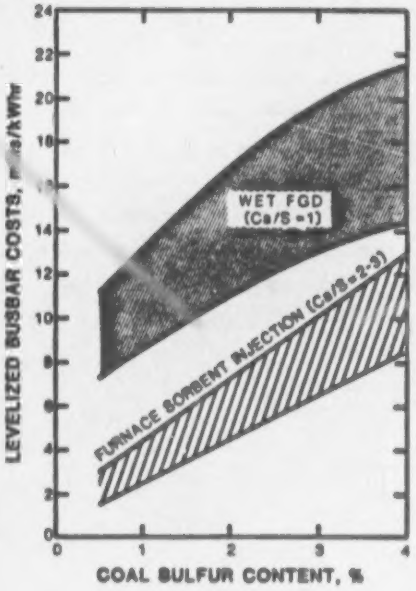


Figure 3-10.

system. For comparison, the approximate capital costs of a wet limestone scrubber are also shown in Figure 3-9. The wet scrubber costs are based on new unit applications that have been adjusted for retrofit. The capital costs of the furnace sorbent injection system offer a capital cost advantage of $100/kW over wet scrubbing systems. This is substantial cost savings reinforcing the potential attractiveness of the technology for incremental SO_2 controls in the range of 40-70 percent.

Operating costs of the furnace injection process are dominated by the cost of the sorbent and the increased waste disposal costs and depend on the sulfur content of the coal and the required sorbent injection rate (calcium-to-sulfur ratio), but again are substantially lower than wet scrubbing. Figure 3-10 shows the increased cost of electricity (30-year levelized busbar costs, mills/kW-hr) as a function of the coal sulfur content. This figure also includes the approximate levelized busbar costs for a retrofit wet scrubber installation. Despite the higher quantity of sorbent required with furnace sorbent injection compared to the wet scrubber, the levelized busbar costs are anticipated to be approximately one-half that of wet scrubbing for an equivalent coal sulfur application. The lower costs for the limestone injection system can, in large part, be attributed to the simplicity of the process, if successful.

3.2 POST FURNACE DRY SORBENT INJECTION FOR SO_2 CONTROL

An alternative dry sorbent injection process involves the injection of a pulverized sorbent downstream of the boiler at the inlet of a fabric filter. The dry sorbent material collects on the surface of the bags which then serve as a chemical contactor where the SO_2 reacts with the sorbent. The resulting sulfate is then removed as a dry product along with the fly ash. Sorbent materials of most interest for this process are nahcolite (naturally occurring sodium bicarbonate, $NaHCO_3$) and trona (naturally occurring sodium sesquicarbonate $NaHCO_3 \cdot Na_2CO_3 \cdot 2H_2O$). Currently, only trona is available commercially. Other sorbent possibilities include ammonia and alkalized aluminum.

EPRI has had an on-going development of post furnace dry SO_2 removal that has recently culminated in a successful full-scale demonstration of the process (RP-491-1, RP-982-8, RP-1682-2, RP-1682-1). This full-scale demonstration was jointly sponsored by EPRI, Public Service Company of Colorado (PSCC) and Multi Mineral Corporation and was carried out at the 22 MW Unit 1 of PSCC's coal-fired Cameo station near Grand Junction, Colorado. During the full-scale demonstration the SO_2 removal efficiency of the dry injection system was evaluated as a function of the sorbent feed rate and the baghouse operating parameters. The results support the findings of a previous laboratory study sponsored under RP 982-8. In the Cameo tests, when stoichiometric amounts of the sorbent were used, an SO_2 removal of 80 percent was achieved with nahcolite and 55 percent removal with trona. Soda ash was ineffective. The SO_2 removal characteristic as a function of the amount of material injected is shown in Figure 3-11. (The sorbent injection rate is shown as a fraction of the amount theoretically required to react with all of the SO_2 in the flue gas to form sodium sulfate, Na/S molar ratio = 2.)

The impact of the dry injection process on the overall operation of the baghouse was minimal. Removing the fly ash-reacted sorbent cake from the fabric was not significantly more difficult than the normal practice of collecting fly ash only. These findings showed that the injection process does not impose a significant pressure drop penalty.

Following the Cameo demonstration project, EPRI initiated an economic evaluation of dry injection for both nahcolite and trona. The results indicate that low capital cost will be an attractive feature of this technology with reagent cost and transportation being the controlling economic factors. Capital costs and 30-year levelized busbar costs have been estimated for a 500 MW unit burning a 0.5 percent sulfur coal, with 70 percent SO_2 removal and a delivered sorbent cost of $100/ton for nahcolite and $75/ton for trona. The capital requirement for the injection system is approximately $22/kW ($90/kW if the fabric filter is included.) The levelized busbar costs are 9.6 and 8.6 mills/kW-hr for nahcolite and trona respectively (including 3.4 mills/kW-hr for operation of the fabric filter). Again, both the capital

POST-FURNACE SO_2 REMOVAL WITH NAHCOLITE AND TRONA

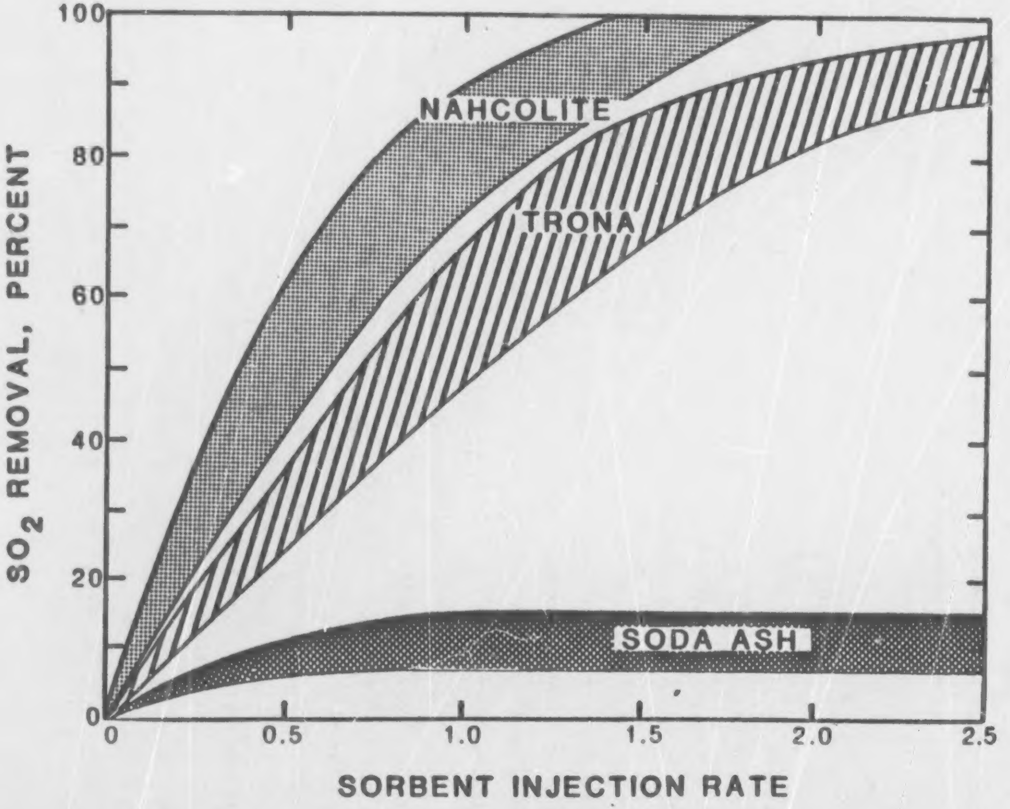


Figure 3-11.

costs and the levelized busbar costs are substantially less than wet scrubbing. Because of the relatively high projected costs for the sorbent (nahcolite and trona) and sorbent transportation, this SO_2 removal technology appears to be most suitable for low sulfur western coal applications if used for full SO_2 control. New sorbent materials under consideration may extend the applications of this technology to eastern regions where current sorbents now appear uneconomical.

SECTION 4.0

CONCLUSIONS

Conclusions drawn from this assessment of retrofit NO_x and SO_2 control are summarized below:

NO_x RETROFIT CONTROL

- Electric utility boilers represent 65 percent of the national SO_2 emissions and 29 percent of the national NO_x emissions. Coal-fired units produce most of these emissions (>80 percent).

- NO_x control by means of combustion modifications have historically been more cost effective than flue gas treatment. Combustion modifications have been used to meet current new unit regulations (69 units are operating with NO_x controls at costs generally less than $5/kW).

- Foreign technology from Japan and W. Germany in the form of low-NO_x burners and innovative combustion methods have produced NO_x levels less than 200 ppm (>50 percent reduction).

- NO_x emissions are directly dependent upon the boiler and burner design, boiler firing practice, and to a much lesser degree, the coal type.

- The two most common boiler design configurations are wall-fired and tangentially fired. Tangentially fired units represent more than 40 percent of the capacity but only 30 percent of the emissions because they are low emitters. Wall-fired units produce NO_x emissions that are disproportionately larger than their installed capacity.

- Cyclone-fired units and wall-fired units equipped with cell burners produce high-NO_x emissions and no retrofit options are commercially available at this time.

- NO_x combustion control options are frequently classified into two major categories: (1) operational modifications and (2) hardware modifications. The NO_x reduction potential is dependent upon the boiler design category to which it is applied.

- -- The two most cost-effective operational modifications are low excess air firing and biased firing. Reductions up to 15 percent are possible at costs of approximately \$0.40/kW.
- -- The installation of low-NO_x burners or overfire air ports are common, but more costly, hardware retrofit modifications. NO_x reductions of 30-70 percent can be achieved at costs ranging from \$2 to \$15/kW.
- -- NO_x reductions and installed costs are very site-specific and variations of 2 to 1 are not uncommon.

- The choice of a NO_x control option is a complex trade-off between (1) capital and operating cost, (2) NO_x reduction potential, (3) physical compatibility between the modification and the furnace design, (4) outage requirements, and (5) site-specific factors.
- Nearly 60 percent of the boiler capacity greater than 100 MW are potential candidates for some form of NO_x retrofit. However, because of design and operational constraints, only 38 percent of the pre-NSPS capacity greater than 100 MW is suitable for retrofit with commercially available low-NO_x burners.
- EPRI has a number of contracts to develop retrofit NO_x combustion controls for wall-fired and tangentially fired units and examine retrofit trade-offs for all boiler design categories.
  - -- KVB is preparing a NO_x retrofit guideline to assist utilities in these decisions.
  - -- CE is evaluating a low-NO_x PM burner for tangentially fired units on U.S. coals. NO_x reductions of 45-60 percent are projected for full-scale retrofits. Limestone injection tests are being performed to assess the SO_2 removal potential under low-NO_x conditions.
  - -- Riley-Stoker is developing retrofit low-NO_x burners and advanced fuel/air staging for wall-fired units. NO_x reductions of up to 60 percent have been achieved using the CCV burner which has a high retrofit potential.

  - Babcock & Wilcox is conducting a similar program for cell burner units. These units represent a substantial fraction of total utility NO_x emissions but no retrofit burner option is commercially available at this time.

SO_2 RETROFIT CONTROL

- Utilities need simple, lower cost alternatives to complex SO_2 scrubbers/spray dryers for retrofit to existing boilers. This is especially true where incremental (< 50 percent) emission reductions are required.

- Dry SO_2 control processes which can be combined with conventional boiler/particulate collection equipment appear most attractive.

  - Post-furnace sodium-based sorbent injection into a fabric filter particulate collector for 80 percent removal on low-sulfur coal.

  - In-furnace calcium-based sorbent injection for 50 percent removal on all coals.

  - Neither dry sorbent injection processes (e.g., in-furnace or post-furnace) are new concepts having been investigated in the 1960's. The criteria under which these processes are being evaluated and applications considered have changed over the ensuing years producing a renewed interest in their further development.

- Historically, work with in-furnace sorbent injection for SO_2 control was not considered a technical success since SO_2 removals were limited to at most 50 percent. This early work lacked a sufficient understanding of the process.

- Recent pilot-scale test results indicate that the thermal environment produced by low-NO_x combustion systems may provide an opportunity to optimize the sorbent injection process with SO_2 removals of 50 percent achievable on a retrofit basis for medium to high sulfur coal (+ 2 percent sulfur).

- Assuming furnace sorbent injection requires retrofit with low-NO_x combustion systems, only about 38 percent of the pre-NSPS boiler capacity appears technically amenable to furnace limestone injection at this time. Current research and development programs for low-NO_x combustion systems may expand this applicability.

- Application of furnace sorbent injection for SO_2 control requires an understanding of the potential impacts on boiler performance. Understanding these issues has been a major part of EPRI's program in this technology area.
  - Boiler particulate loadings can be doubled for a typical application of the technology (2 percent sulfur and 10 percent ash coal, and a sorbent injection rate corresponding to Ca/S = 2). This can impact deposition, convective section erosion and ash handling.
  - Dry sorbent injection can alter ash chemistry impacting slagging and fouling characteristics. Pilot-scale studies indicate that deposition rates will increase in both the radiant and convective section but the deposits are more easily cleaned. Further study is warranted to assess deposition impacts in the air heater.
  - The increased loadings and change in chemical composition can impact the performance of an electrostatic precipitator (ESP). Further work is warranted to establish this impact and determine whether upgrading of the ESP or retrofit is necessary.
- Expanded research and development programs to resolve these issues are underway in the U.S. by EPRI, individual utilities and EPA for high sulfur coal and in West Germany for low sulfur coal. This is expected to lead to prototype testing in the U.S. on a high sulfur, 50-100 MW_e utility boiler by 1986 and a large, 400 MW_e demonstration by 1988.
- Capital costs of furnace limestone injection are expected to range from \$15 to \$50/KW_e plus any increased ash collection, handling and disposal costs. This is about 10 to 30 percent of the equivalent capital cost of a lime/limestone scrubbing system.
- This capital cost advantage of furnace limestone injection, however, is partially offset by the increased amount of limestone required. This means that the increase in cost of electricity is about one-half that resulting from use of a lime/limestone scrubber. This does not include consideration of boiler reliability impacts which are not yet definable for the furnace limestone injection process.
- Post-furnace injection of a sodium-based dry sorbent ahead of a fabric filter has been demonstrated at full scale to produce SO_2 removals in excess of 70 percent at reasonable sorbent injection rates for low sulfur western coals.

- The estimated costs of post-furnace sorbent injection are substantially lower than wet FGD system for low sulfur western coals (0.5 percent sulfur); capital costs are estimated to be 22 \$/kW (90 \$/kW including the fabric filter) and levelized busbar costs are 9.6 to 8.6 mills/kWhr depending on the specific sorbent (including 3.4 mills/kWhr for operation of the fabric filter).

R&D NEEDS

NO_x:

-- Assess the national NO_x emissions reduction potential

-- Develop current information on cost, performance, and applicability of available NO_x retrofit options

-- Develop new low-NO_x combustion technology for retrofit applications

SO_x:

-- Develop process information to optimize SO_2 removals on a site-specific basis

-- Assess the potential impacts of furnace sorbent injection on boiler system operation

This Electric Power Research Institute paper, presented at the Joint Symposium on Stationary NOx control in Dallas, November 1982, describes the current status of three important dry processes to simultaneously control SOx and NOx: copper oxide, carbon absorption and election-beam irradiation. Sulfur oxide removal efficiencies exceed 90 percent in each case, while nitrogen oxide removal exceeds 70 percent. None of the processes are commercially available yet but development is proceeding in the U.S., Japan and West Germany.

EPRI

CURRENT STATUS OF DRY NO_X-SO_X EMISSION CONTROL PROCESSES

S. M. Dalton

Presented to

1982 Joint Symposium on Stationary NO_x Control

Dallas, Texas
November 1-4, 1982

ELECTRIC POWER RESEARCH INSTITUTE

CURRENT STATUS OF DRY NO_x-SO_x EMISSION CONTROL PROCESSES

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ABSTRACT

Combined NO_x-SO_x control techniques using a variety of complex chemical flue gas treatment processes have been proposed by a number of manufacturers. The wet chemical NO_x-SO_x processes are generally not economically competitive and development has shifted to dry NO_x-SO_x techniques. Three dry techniques that are currently being widely pursued as alternatives to wet FGD/selective catalytic reduction are discussed in some detail in this paper: the copper oxide processes, the carbon adsorption processes, and the electron-beam (E-beam) irradiation processes. There are at least five versions of the copper oxide process, eight versions of the carbon adsorption process with NO_x-SO_x reduction and two versions of the E-beam concept. All of these processes appear capable of achieving over 90% SO_2 removal and greater than 70% NO_x removal, although the power plant economic and operational impacts are not fully understood. None of the processes are in service commercially, though active development is underway in the U.S., Japan, and Germany.

NO_x-SO_x PROCESS STATUS

A number of simultaneous NO_x-SO_x removal processes have been developed in the past two decades. Early work included wet and dry scrubbing processes based on a variety of chemical reactions and unit operations. Today, dry combined processes that remove SO_x and NO_x simultaneously are being developed to compete with SO_x and NO_x removal done as independent steps. Three of the most advanced types of systems for combined dry SO_x-NO_x removal are copper oxide, carbon adsorption and electron-beam irradiation. This paper will discuss the status of systems using these three principals and will discuss one example of each system type in more detail. The developmental status, process description, performance data and relative costs will be summarized for each process example.

HISTORY OF NO_x-SO_x

Much of the early work in combined NO_x-SO_x processes evolved out of the initial work in wet scrubbing. Many early wet scrubbers for SO_2 were modified to try to sorb NO_x-usually as equal molar proportions of NO and NO_2. Unfortunately for wet scrubbers, the NO_x emissions from utility boilers are predominantly NO and in order to absorb have to be oxidized to reach equal molar proportions of NO and NO_2. This requiresozone, EDTA, $FeSO_4$, or other means to allow adsorption or to oxidize the NO. In addition, the co-absorption of NO and NO_2 is a slow process and required large absorbers with high pressure drop and many stages. When this was added to the complex chemistry of the process requiring many steps, the results were processes that could not compete economically with independent SO_x and NO_x control steps (such as conventional SO_x scrubbing plus catalytic NO_x removal).

A few systems were not limited by the chemistry and mass transfer limitations of wet scrubbers. All three concepts that are discussed in more detail in this paper do not suffer from this limitation. The dry copper oxide and carbon adsorption concepts have been proposed for NO_x and SO_x control by utilities for over ten years, and recent R&D work is bringing them closer to commercialization. The electron beam concept has been under development for a similar amount of time, though the proposed use on coal-fired utility boilers is recent.

GENERIC DESCRIPTION OF NO_x-SO_x CONCEPTS

Copper Oxide

The first generic process is copper oxide. Five systems are based on this generic process. The copper oxide processes depends on formation of copper sulfate to remove SO_2 followed by regeneration of sulfur (using reducing gas) and subsequent re-oxidation of the copper to copper oxide which is the active sorbent form. NO_x is catalytically reduced to N_2. The Shell flue gas treatment process (SFGT) and other several processes are based on the use of copper oxide as shown in Table 1, Reference 1. These include the Nippon Oil Company process (Japanese patent 7729469, March 5, 1977) and the U.S. Bureau of Mines Process, both of which use chemistry similar to the SFGT copper oxide process but are not as well developed. The Exxon Research Engineering processes uses a bed of CuO and Al_2O_3 to form sulfates. NO_x is reduced to N_2 in the presence of copper sulfate and ammonia. Though two U.S. Patents have been issued (4,001,376 and 4,039,478) Exxon has discontinued further development of this process.[1] The Houdry Chemicals Division of Air Products developed a fixed-bed mixture of copper and silica oxides (plus additive) that sorbed gas at 300oC (600oF) and regenerated H_2S using hydrogen. Houdry discontinued work in 1970. The Shell flue gas treatment (SFGT) process is based on copper oxide and is discussed in more detail later in this paper.

Carbon Adsorption

The second generic concept for NO_x-SO_x removal is use of carbon adsorption. At least eight carbon processes claim combined NO_x-SO_x removal as shown in Table 2. This is done via adsorption of SO_x on carbon followed by carbon cleaning (heating, washing, etc.). Carbon acts as a catalyst or has catalyst added so that ammonia addition promotes NO_x reduction to N_2. Carbon pores also catalize oxidation of SO_2 to SO_3 and react with water to form H_2SO_4 which is held in the pores until regeneration. This property allows high carbon loading with acid and reasonable system size.

The Bergbau-Forschung process received the most early attention in the U.S. primarily for SO_x control. It went througn prototype testing in the U.S. at the Scholz plant of Gulf Power in the mid-1970s and at STEAG's station in Lunen, Germany in the mid-to-late 1970s. The first commercial plant is under construction in Mannhein West Germany. Most of the early work was centered around SO_x removal and only lab and pilot work has extended to NO_x removal. The process uses a moving bed adsorber with NH_3 injected upstream for NO_x removal. Hot sand is used to drive SO_2 off of the loaded carbon.

The Chemibau-Reinluft process uses a moving bed similar to the Bergbau-Forschung adsorber. The regeneration is by hot (385°C [750°F]) inert gas. Prototype testing in the 1960s has not led to commercial units.

In Japan, several manufacturers are working on carbon-type DESOX/DENOX processes. The Hitachi process adsorbs SO_2 in a dry carbon bed then washes H_2SO_4 out of the carbon with water. The dilute acid produced (10-20% H_2SO_4) is either concentrated or used to produce gypsum. Test units in Japan were from 2 MW to 150 MW on Tokyo Electric's oil-fired Goi and Kashima Stations. No data on commercial installations has been published in the U.S.

The Sumitomo process uses a moving bed adsorber similar to Bergbau-Forschung and regenerates with hot gas. SO_2 is converted to H_2SO_4 in an acid plant. Sumitomo began work in 1973 with bench work in 1974, and pilot plant work in 1976 on a 3.3 MW equivalent pilot plant. A test at the Takehara station reported results of 90-95% SO_2 removal and 30% NO_x removal at a space velocity of 500-900 hr^{-1}. Later results are not yet published. A demonstration plant with capacity to treat 300,000 Nm^3/hr will be constructed at Matsushima station in 1984(2).

The Takeda chemical process uses an adsorption step similar to the Bergbau-Forschung process and can either use washing or hot gas regeneration. Most recent tests have been done on a 33 MW equivalent plant treating gas from a sintering furnace.

The Unitika process uses fixed reactor beds instead of moving beds. The fixed beds are alternatively regenerated using hot inert gas. Most recent pilot tests have been conducted at the Union Glass Industry plant in Hirakata, Japan.

The only work done entirely in the U.S. is on the Westvaco activated carbon process. This uses a completely different mechanical design of the adsorption step—a multistage fluid bed. The regeneration step uses hydrogen to react with heated char and form H_2S and elemental sulfur at 524°C (1000°F). Pilot plant tests have been conducted on an oil-fired boiler, giving greater than 90% SO_2 removal. Though not tested, NO_x removal should be possible similar to other carbon processes. The process is not being actively commercialized.

The Mitsui-BF Dry DESOX-DENOX process is discussed in much greater detail in the section titled Mitsui-BF dry DESOX-DENOX process. It is an advanced example of the cabon adsorption dry DESOX-DENOX type systems.

Electron Beam Irradiation

The third generic concept that is receiving wide development attention is the electron beam (E-beam) irradiation concept. Two systems using electron beam technology both irradiate the gas using high energy electron "gun" accelerators to dissociate the gases to form free radicals. The two systems being developed differ in end-product, method of humidifying the gas and use of ammonia.(3) In the Avco-Ebara process, flue gas is humidified and cooled to about 79oC (200oF). The gas is irradiated in special reaction chambers by high energy electron beams in the presence of ammonia. SO_x and NO_x are then converted to ammonium sulfate and ammonium sulfate-nitrate. Particles are then collected, agglomerated and sized for eventual use as fertilizer. In the Research Cottrell process, a lime spray dryer is used to humidify the gas and remove some of the SO_x. This is followed by an E-beam chamber optimized for NO_x removal. Since no ammonia is used, the product is calcium sulfite and sulfate with some nitrates and it is not recovered. A fabric filter (or ESP) catches the spent by-product for disposal. Both systems are still at the pilot stage with a 1 ½ MW Research Cottrell Avco-Ebara pilot being built at TVA and a 10 MW pilot being planned in Indiana. The Avco-Ebara system is discussed in detail in the section titled "Avco-Ebara Electron-Beam Process as an example of combined NO_x-SO_x E-beam processes.

SHELL FLUE GAS TREATMENT (SFGT) PROCESS

History and Process Description

The basic technology for the SFGT was developed by Shell Internationale Petroleum Maatschappij B.V. (SIPM) Den Haag, the Netherlands.(4) They began development in 1962. Laboratory and modeling work have been done to predict process performance. Early work was done for oil-derived flue gas. Pilot testing was done at Tampa Electric Company's Big Bend station on a 0.6 MW pilot plant using gas with and without fly ash. The gas was reheated to the proper reaction temperature. This work focused on mechanical and emission control performance. Projections of the acceptor/catalyst fouling rates and life were made. Full results of this testing are discussed in EPA's paper given at an earlier APCA meeting.

Experience on oil fuel is more extensive. A commercial system for combined NO_x-SO_x removal was installed at Showa Yokkaichi Sekiyu (SYS) refinery in Japan. The 40 MW equivalent system, installed in 1973, operates at 90% SO_2 removal and 40% NO_x reduction(1). The refinery already had reducing gas and Claus plant capability so the SFGT system was connected to those unit operations.

The SFGT process uses a fixed bed of copper oxide to sorb SO_2 from flue gas. A flow schematic is shown in Figure 1. Flue gas is passed through a set of specially designed, parallel passage, fixed-bed reactors which use a copper oxide on alumina sorbent for SO_2 removal. The reaction proceeds according to the following equation:

$$CuO(s) + SO_2(g) + 1/2O_2(g) \rightarrow CuSO_4(s) \qquad (1)$$

The reactor configuration was specifically designed to maintain catalyst stability and minimize pressure drop. A cyclical mode of operation using multiple reactors is employed. The system operates at about 385oC (750oF) in both sorption and regeneration steps.

When the majority of the copper in the reactor has been converted to copper sulfate, the flue gas is rerouted to another vessel and the spent vessel is purged in preparation for sorbent regeneration. The sorbent is regenerated by passing a dilute-reducing gas through the vessel.

Either hydrogen or methane can be used as the reducing agent. When methane is used, the regeneration reaction is:

$$CuSO_4 + 1/2CH_4(g) \rightarrow Cu(s) + SO_2(g) + 1/2CO_2(g) + H_2O(g) \qquad (2)$$

If hydrogen is used, reduction occurs by a variety of reactions:

$$CuSO_4(s) + 2H_2(g) \rightarrow Cu(s) + SO_2(g)\ 2H_2O(g) \qquad (3)$$

$$CuSO_4(s) + 4H_2(g) \rightarrow CuS(s) + 4H_2O(g) \qquad (4)$$

$$CuS_4(s) + H_2(g) \rightarrow Cu(s)\ 2H_2O(g) \qquad (5)$$

Sulfide formation is promoted by low temperature in the reactor. After regeneration, the reactor is purged and returned to flue gas service. In previous testing, cycle times of 40 to 80 minutes were measured.

There are several design options for concentrating and preparing the SO_2-regeneration gases for further processing. These options include water absorption/stripping (as shown in Figure 1), solvent adsorption, compression/condensation of water vapor, and complete liquefaction.

Conversion of the concentrated SO_2 to elemental sulfur can be accomplished by several technologies including the Allied Chemical, RESOX, BAMAG, or Claus processes. Most of these technologies also require a step to make reducing gas. Options for reducing gas include steam-methane or steam-naptha reforming and coal gasification.(1)

The CuO and $CuSO_4$ formed during oxidation and SO_2 acceptance act as catalysts for the reduction of NO_x by NH_3 addition. The acceptor is noncatalytic in the form of Cu. All reactions shown below are exothermic and raise flue gas temperature of 3oC (5oF)(4)

NO_x and NH_3 Reactions

$$6NO + 4NH_3 \rightarrow 5N_2 + 6H_2O \qquad (6)$$

$$4NO + 4NH_3 + O_2 \rightarrow 4N_2 + 6H_2O \qquad (7)$$

$$4NH_2 + 3O_2 \rightarrow 2N_2 + 6H_2O \qquad (8)$$

Shell Flue Gas Treatment Process Economics

The (SFGT) economics have been reviewed recently in a study by Stearns-Roger for EPRI.(4) The cost estimates in that report were based on mid-1978 dollars for a mid-1979 operating date. All costs in the summary cases in tables 3 and 4 were escalated to December 1982 dollars using EPRI's Technical Assessment Guide (TAG) economic premises. Table 3 shows capital cost estimates for various levels of NO_x control and Table 4 shows first year operating costs.

Process Limitations

The technical limits of the technology are:

- The process includes chemical processing steps that are more typical of a refinery--requiring additional operator training.
- Deposits in downstream ductwork and airheaters may require redesign of equipment.
- Excess moisture from tube leaks, condensation, or sootblowers can dissolve the soluble copper oxide/copper sulfate catalyst.
- Reducing gas is required necessitating reformer steps.
- Control and fire prevention methods will need special care during design and operation.
- The system is integrated with power plant heat recovery equipment and the SFGT cycle may be upset during power plant equipment malfunctions.
- Surge capacity is limited between SFGT acceptor/regeneration and Claus.
- NH_3 carry-over guarantees are not being made by UOP due to difficulty of measurement.

These technical limits are probably overshadowed by the regulatory requirements for NO_x control. Unless post-combustion NO_x control is required, the economics and complexity will probably not mean wide use of the SFGT process on utility boilers.

MITSUI-BF DRY DESOX/DENOX PROCESS

History and Process Description

Early development of the Mitsui-BF dry DESOX/DENOX process grew out of the work done by Bergbau-Forschung in Germany. The recent Japanese work included laboratory tests in 1980-1981 followed by absorber tests starting in early 1981.(5) A prototype by-product pilot plant was constructed in late 1981 and early 1982.

The Mitsui-BF process is divided into an adsorption section, a desorption section and a sulfur recovery section. The adsorber concept is similar to the Bergbau-Forschung with carbon regenerated by indirect heating. The rich SO_2 stream produced is partially reduced in a Mitsui-designed reduction reactor and the H_2S/SO_2 mixture is catalytically converted in a Claus reactor to elemental sulfur. This process is illustrated in Figure 2.

Adsorption/Regeneration Section. Flue gas enters the moving bed adsorbing system which holds activated coke. Since the optimum flue gas temperature for the process ranges from 120 to 140oC (248 to 284oF), which is usual with flue gas behind air-preheater, the process does not require preheating or precooling. The adsorption system consists of two vertically connected adsorbers where activated coke pellets flow downward by gravity from the top adsorber (second stage) to the bottom adsorber (first stage). Flue gas flows horizontally through the bottom adsorber (first stage) and goes into the top adsorber (second stage) via a duct. The first and second stage adsorber share the duty of SO_x removal and NO_x removal and both are fed ammonia. Chemical reactions in the adsorption section are as follows:

SO_2 removal

$$2SO_2 + O_2 + 2H_2O \rightarrow 2H_2SO_4 \qquad (9)$$

NO removal

$$4NO + O_2 + 4NH_3 \rightarrow 4N_2 + 6H_2O \qquad (10)$$

Competing ammonia reactions

$$NH_3 + H_2SO_4 \rightarrow NH_4HSO_4 \qquad (11)$$

$$2NH_3 + H_2SO_4 \rightarrow (NH_4)_2SO_4 \qquad (12)$$

Competing ammonia reactions 11 and 12 are favored kinetically and therefore ammonia injection with high SO_x levels results in lowered NO_x removal. The best NO_x removal occurs at low SO_x levels when the proper amount of ammonia is available to react with NO_x in contact with the coke. This is the reason the Mitsui-BF process uses two reactors. The lightly loaded coke that is most active for NO_x removal is in the second stage, where sulfuric acid, ammonium sulfate and bisulfate have not built up. As they become loaded, they are sent to the first stage for SO_x removal duty, where there is still enough activity to remove SO_x.

When high NO_x removal efficiency is required, it is necessary to increase ammonia feed to the second stage and consequently ammonia residual appears behind the adsorber. Test results show residual does not exceed 20 ppm at a NO_x removal efficiency of more than 80 percent. However, if treatment of the residual is required, the flue gas is sent through an extra adsorber which is smaller than the former ones and contains sulfuric acid rich activated coke from the first stage, to remove ammonia according to the rapid competing ammonia reactions.

Dust load in the flue gas behind these adsorbers is less than 30 mg/Nm^3-dry (.014 gr/scf) and it may be possible to send the flue gas to the stack without dust collectors. In the Mitsui pilot plant, a bag filter is provided in order to provide backup particulate control, and outlet dust load is less than 1 mg/Nm^3-dry (.5 x 10^{-3} gr/scf).

Activated coke saturated with sulfuric acid is discharged from the first stage adsorber by vibrating chute and sent to the desorber for regeneration. Activated coke is regenerated by indirect heating. The temperature of activated coke in the heat exchanger is maintained at (400±20°C) (752±36°F). As a result of regeneration, concentrated gas with 20% SO_2 by volume is discharged from the desorber. Chemical reactions in the desorption section are as follows:

$$H_2SO_4 \rightarrow H_2O + SO_3 \qquad (13)$$

$$NH_4HSO_4 \rightarrow NH_3 + H_2O + SO_3 \qquad (14)$$

$$2SO_3 + C \rightarrow 2SO_2 + CO_2 \qquad (15)$$

$$3SO_3 + 2NH_3 \rightarrow N_2 + 3SO_2 + 3H_2O \qquad (16)$$

The regenerated activated coke is indirectly cooled by air and recycled to the adsorber.

Sulfur Recovery Section. As a tail-end process for SO_2 enriched gas treatment, the Mitsui pilot plant is provided with a sulfur recovery process. The process consists of conversion of some of the SO_2 to H_2S and subsequent Claus reaction. Two-thirds of the SO_2-enriched gas enters the coke reactor, and SO_2 is reduced to H_2S with steam and hot coke (coke breeze is used due to its low cost and availability). The temperature of the coke reactor is maintained between 650 to 950°C (1202 to 1742°F). H_2S from the coke reactor is mixed with the by-passed one-third of SO_2 gas and sent to subsequent Claus reactor. Many reactions occur in the coke reactor, but representative reactions can be written as follows:

$$C + SO_2 \rightarrow S + CO_2 \qquad (17)$$

$$C + H_2O \rightarrow H_2 + CO \qquad (18)$$

$$H_2 + S \rightarrow H_2S \qquad (19)$$

Claus reaction (in Claus stages):

$$2H_2S + SO_2 \rightarrow 3S + 2H_2O \qquad (20)$$

The tail gas from the Claus unit enters an incinerator to convert residual sulfur and H_2S to SO_2 and goes to the adsorber inlet or stack. To maintain high SO_2 removal efficiency without increasing running cost as a total system, the sulfur recovery process has to perform at high sulfur conversion efficiency. This is the reason Mitsui chose to develop the sulfur recovery process consisting of a coke reactor and Claus unit instead of more conventional processes.

Test Results

1. Typical NO_x-SO_x removal results are summarized in Table 5. Ammonia was injected to both first and second stages and the total molar amount did not exceed two times the inlet NO_x throughout these tests.

2. Figure 3 shows the effect of flue gas temperature on SO_2 and NO_x removal efficiency in the first stage. As flue gas temperature becomes higher, SO_2 removal efficiency decreases rapidly while NO_x removal efficiency increases slightly. NO_x removal efficiency in the first stage is also a function of ammonia feed, but this effect is not as significant as the temperature effect.

3. Figure 4 shows the relation between NO_x removal efficiency and flow rate of activated coke in the second stage. Since activated coke flows through both adsorbers in series, good NO_x removal efficiency in the second stage as shown in Figure 4 depends on good SO_2 removal efficiency in the first stage. This is because high SO_2 removal efficiency in the first stage is required to avoid competition for ammonia by SO_2.

A process material and energy balance is shown in Figure 5. The gas flow represents approximately a 350 MW plant with medium sulfur coals.

Mitsui-BF Process Economics

The Mitsui-BF dry DESOX-DENOX process represents an advanced design that has not yet been built at full scale. The developer, Mitsui, has provided EPRI with their estimate of cost for a DESOX-DENOX system on a 350 MWe (nominal) coal-fired power plant with the same material and energy balance as was shown in Figure 5. The basis of this estimate is not identical to others in this paper since it does not include owner costs such as land, interest during construction and project and process contingency, etc., in the same manner as the EPRI estimates. For example, EPRI's standard estimates usually show a higher ratio of indirect costs to direct cost. Reported initial capital charges for a 350 MW plant are:

	Million Dollars*	Million Yen
Direct (Including Initial Coke Charge)	37.5	9,932
Indirect	3.6	950
Total	41.1	10,882

Annual Operating Costs (First Year)

	Million Dollars	Million Yen
1. Fixed Costs	9.09	2411
2. Operating & Maint. Costs		
Labor	.20	54
Repair	1.23	325
Utility	5.37	1423
Other	.41	109
3. By-Product		
Sulfur Credit	(.37)	(98)
4. Overhead and Tax	.96	254
TOTAL First Year Costs	16.89	4,479

*Note conversion rate is 265.1 Yen/$ - costs not according to EPRI premises.

The technical limits of the carbon adsorption process are:

- Reduced NO_x removal with very high SO_2 inlet
- Potential for plugging of adsorber with ammonium salts
- Need to minimize ammonia breakthrough and carbon loss
- Reactor temperature control/hot spot and fire control
- Commercial adsorbent availability and cost
- Adsorbent attrition and consumption rate

The approach used by Mitsui (splitting the absorber bed functions) is an example of an approach being used to refine this well-established process concept.

AVCO-EBARA ELECTRON-BEAM PROCESS

History and Processes Description

Early development (from 1970) of the irradiation concept was done in Japan by Ebara in combination with the Japan Atomic Energy Research Institute(3). The original Ebara process concept evolved from a consideration of the reactions in photochemical smog. Batch-type tests using relatively low energy irradiation and long exposure times proved the formation of an aerosol was possible. From 1972-1974, small flow-type irradiation devices were used with a more powerful accelerator. The removal of SO_x and NO_x was good at the high irradiation rates and improved at lower temperature. From May 1974 to

February 1977, a pilot plant with 1000 Nm^3/hr (625 scfm) was constructed and operated by Ebara. In the first half of the testing, unstable solid particles or a mixture of sulfuric and nitric acids was collected by an ESP. In the second half of the tests ammonia addition yielded a stable dry powdered product and improved SO_x and NO_x removal. The final stage of Japanese development included a 10,000 Nm^3/hr (6250 scfm) pilot plant built at an iron works in Japan and applied to sintering exhaust gas for one year (1977-1978) normal conditions were 200 ppm SO_x and 180 ppm NO_x, respectively. During this time, reported SO_2 removal efficiency exceeded 90 percent and NO_x removal exceeded 80 percent at 1.0 Mrad (equivalent to 10 J/g of gas or 4.54 KW-sec/lb). To maintain high efficiency, electron dose rates had to be uniform. Results of the large-scale Japanese test work are summarized in Figure 6 and 7 which show the SO_x and NO_x removal as a function of reaction temperature and absorbed energy.

Avco, the licensor and co-developer of this process in the U .S., has been actively pursuing a demonstration project in this country.[3] The process demonstration unit is expected to have a capacity of 32,000 Nm^3/hr or 20,000 scfm (about equivalent to 10 MWe gas flow). Avco has done laboratory work on the concept and has investigated U.S. markets for the by-product. The proposed full-scale demonstration has been supported by DOE and a group of Indiana utilities as well as the state of Indiana. Ebara is reportedly taking over U.S. development initiative and actively seeking funding.

The process steps include humidification and temperature control, ammonia addition, irradiation and by-product collection. This is shown schematically in Figure 8. The humidification vessel uses sprays to reduce bulk gas temperature to 79°C (200°F). Ammonia is added to make the by-product usable as ammonium sulfate and nitrate. The irradiation produces ozone and free radical intermediates, such as N, O, OH, HO_2. These combine to give sulfuric and nitric acids and ammonium salts. The overall process reactions can be summarized as:

$$SO_2 + NO + NH_3\ H_2O \xrightarrow{O,\ OH,\ HO_2} (NH_4)_2\ SO_4 + (NH_4)_2\ SO_4 + 2NH_4NO_3 \qquad (21)$$

Water is required to give proper radical formation necessary for SO_x removal.

For simultaneous SO_x and NO_x removal the optimum process conditions are different. High NO_x removal (>90%) requires lower reaction temperature.

The electron beam irradiation step is at the heart of the design. Key design parameters are irradiation dosage, electron beam penetration, electron beam efficiency (absorbed energy vs. electrical power required), gas mixing, and reaction temperature. The reaction chamber designed to do this includes high voltage electron guns that discharge a beam of electrons into the gas through air cooled, thin titanium windows and sweeps the gas volume at a rate of 200 Hz. About 800 kW appears to be a good design value for the electron beam voltage because it will allow cable connections.(3) A detailed discussion of the process design and control variables is contained in references 3 and 6.

Avco-Ebara Process Economics

The economics of the Avco-Ebara process have been evaluated for EPRI by Bechtel Group, Inc., of San Francisco in Reference 3. This evaluation is based on December 1980 pricing and standard premises specified by EPRI. The costs have been escalated to December 1982 for presentation in this paper.

The capital costs are shown in Table 1 and first years costs in Table 7 along with unit rates for labor, utilities, and chemicals. Details are presented in Reference 3. Total levelized revenue requirement is 32.9 mills/kWh assuming no by-product credit. The study also prepared a comparison between conventional limestone and the Avco-Ebara process. This is shown below in December 1982 dollars.

Process	Total Capital Requirement ($/kW)	First-Year Operating Cost ($1000)	30-Year Levelized Revenue Requirement (mills/kWh)
Avco-Ebara	333.0	57,332	32.9
Wet Limestone	212.5	38,503	21.9

On the basis of the present state of development for, and the extent of experience with, the Avco-Ebara process and conventional wet limestone scrubbing, the levelized revenue requirement for the Avco-Ebara process without by-product credit is as much as 1.5 times greater than the levelized revenue requirement for conventional wet limestone scrubbing. In situations where postcombustion NO_x control is not required, the Avco-Ebara process becomes economically competitive with conventional wet limestone scrubbing only if a 58 $/ton net-back for the by-product can be obtained. In situations where postcombustion NO_x control is required, Bechtel estimates that the Avco-Ebara process, without by-product credit, may provide up to a 25 percent savings in the levelized revenue requirement over a combination of wet limestone scrubbing plus selective catalytic reduction.

Both versions of the electron beam technology represents an approach to combined NO_x-SO_x control based on novel physical and chemical steps. Technical limits of the technology are:

- For utility flue gas treatment, electron beam systems more powerful than present production models are required (High Voltage Engineering Corporation of Burlington, Massachussetts probably could supply these now);
- Proper control of humidification is required to prevent corrosion;
- The salability of by-products with and without fly ash will be critical to process economics;
- Coal-fired power plant flue gas has not been tested in the U.S., and a prototype is a necessary next step in development;
- A high ratio of electrical energy (versus steam or fuel) is required to supply auxiliary power needs.

The novel nature of this process means a prototype unit demonstration is an essential next step before commercial units are purchased.

CONCLUSIONS

All three dry generic NO_x-SO_x treatment processes discussed in this paper represent advanced methods of emission control.

All use ammonia and a common problem is ammonia monitoring, breakthrough and handling.

By-products and unit operations are not typical of utility plants. These by-products will require some special training and design considerations.

None of the processes have been tested at full-scale on coal-fired plants and commercial acceptance usually requires developers to complete this step before widespread use.

These processes are more complex mechanically than conventional wet scrubbing for SO_2, but may make inroads in the emission control market if more stringent NO_x control is required. Out of the many processes that have been proposed for dry, combined SO_x and NO_x control, these three generic concepts stand out as likely candidates for consideration on future plants.

ACKNOWLEDGEMENTS

I would like to acknowledge the efforts by Mssrs. Kato and Kitamura of Mitsui in providing Mitsui-BF dry DESOX/DENOX process data for this paper. Additionally I would like to acknowledge the use of information in three draft reports, one prepared by Bechtel Energy, Inc. on the Avco-Ebara Process, one by Stearns-Roger Engineering on Postcombustion NO_x Processes, and one by Radian Corp. on the Status of FGD Processes. Information from these drafts was helpful in preparing this paper.

I would also like to thank Mssrs. Cichanowicz and Morasky of EPRI for their assistance in preparing the paper and reviewing its accuracy.

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1. Radian Corporation, The Evaluation and Status of Flue Gas Desulfurization Systems (Draft); Palo Alto, RP982-28, June 1982.

2. Ando, J., SO_x and NO_x Removal for Coal-Fired Boilers in Japan; preprints of EPA/EPRI Flue Gas Desulfurization Symposium, Hollywood, Florida, May 1982.

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4. Stearns-Roger Engineering Corporation, Technical and Economic Feasibility of Ammonia Based Postcombustion NO_x Control; Palo Alto, California, CS-2713, November 1982.

5. Personal Communication with Mssrs. Kato and Kitamura, Mitsui Mining and Mitsui and Co., USA, September 14, 1982.

6. Shui, V. H., and Carlson, D. G., Conversion of SO_x and NO_x to Fertilizer by Electron Irradiation; AICHE National Meeting, Cleveland, Ohio, Paper No. 56F, August-September 1982.

TABLE 1. COPPER OXIDE SORPTION SYSTEMS

Process	Type[a]	Status[b]	SO_2 Removal Capability[c]	Process Complexity[c]	NO_x Removal[d]	Particulate Removal	Clean Gas Required[d]	Subsystem Used[c]
BOM Copper Oxide	DR	DOE pursuing E	M	M	?	N	Y	R
Exxon Copper Oxide	DR	CP	M	M	Y	N	Y	R
Houdry	DR	E	M	M	?	N	Y	R
Nippon Oil Copper Oxide	DR	?	?	?	?	?	?	?
Shell Copper Oxide	DR	P	M	M	Y	N	Y	R

Source: Ref. 1--Radian

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a) W = Wet, D = dry, T = throwaway, R = regenerable.

b) C = commercial, CU = commercially used in other industries, CA = commercially available.
P = prototype, E = experimental CP = conceptual

c) H = high (95%), M = medium (80-95%), L = Low (<80%) for SO_2 removal.

d) Y = yes, N = no.

e) S = sulfur gas treatment, R = reducing gas production, W = waste and purge stream treatment, N = no.

TABLE 2. CARBON BASED SORPTION SYSTEMS

Process	Status[a]	SO_2 Removal Capability[b]	Commercial Technology Used[c]	Particulate Removal[c]	Subsystem Used[d]
Bergbau-Forschung	CA	H	Y	Y	R
Chemibau-Reinluft	P	H	Y	Y	R
Hitachi	P	H	Y	Y	N
Mitsui	P	H	Y	Y	S
Sumitomo	P	H	Y	N	S
Takeda	P	H	Y	Y	S
Unitka	E	H	Y	Y	S
Westvaco	E	H	N	Y	S

Reference 1-Radian

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a) C = commercial, CU = commercially used in other industries, CA = commercially available.
P = prototype, E = experimental CP = conceptual

b) H = high (95%), M = medium (80-95%), L = Low (<80%) for SO_2 removal.

c) Y = yes, N = no.

d) S = sulfur gas treatment, R = reducing gas production, W = waste and purge stream treatment, N = no.

TABLE 3. CAPITAL COST ESTIMATE SHELL FGT PROCESS
($\$10^3$, Dec. 1982)

Case		1	2	3
NO_x Reduction %		67	83	90
Inlet NO_x, ppm		300	300	300
Reactors		2884	3171	3171
Ductwork		1827	1827	1827
Insulation		604	604	604
Ammonia Storage		423	582	503
UOP Package		8079	9105	9196
Flow Smoothing		1706	1706	1706
Claus Package		2265	2265	2265
Reformer Package		1691	1691	1691
Differential Costs		3624	3624	3624
Other Direct Costs		8101	8350	8365
Other Materials		1513	1561	1561
Field Labor		13450	13874	13878
Indirect Costs		8574	3843	18850
Process Capital		54741	57103	57241
(Cost $/kW)		(109)	(114)	(114)
General Facilities		2938	2860	2863
Engineering & Fees	o A/E	5764	5764	5764
	o UOP	17078	17078	17078
Contingency	o Project	12590	3154	13167
	o Process	5474	5718	5724
Sales Tax		1137	1199	1200
Total Plant Investment		99522	102876	103037
(Cost $/kW)		(199)	(206)	(206)
Royalty Allowance		227	257	257
Preproduction		5646	5871	5880
Inventory		486	524	529
Initial Catalyst		8003	9611	9611
Land		0	0	0
Total Capital Requirement		113,884	119,139	119,314
(Cost $/kW)		(227)	(238)	(238)

TABLE 4. FIRST YEAR O&M COSTS FOR 500 MW SHELL FGT PROCESS
($\$10^3$, Dec. 1982)

Case	1	2	3
NO_x Reduction %	67	83	90
Inlet NO_x, ppm	300	300	300
Fixed O&M			
Operations	331	331	331
Maintenance	1616	1685	1685
O&M Overhead	977	1004	1004
TOTAL Fixed O&M	2924	3020	3020
Variable O&M			
Ammonia	498	619	664
Steam	80	83	83
Power	2718	3005	3005
Naphtha	1963	1963	1963
Sulfur Credit	(196)	(196)	(196)
Catalyst/Acceptor	8003	9611	9611
Reheat Credit	(1157)	(1157)	(1157)
TOTAL Varible O&M	14,223	16,242	16,287
Levelized Cost Mills/kWh	20.3	22.2	22.3

TABLE 5. TYPICAL PERFORMANCE DATA IN MITSUI-BF DRY DESOX/DENOX PILOT PLANT

Inlet Gas Condition									
Flue gas handled	Nm^3/hr (SCFM)	1000	(622)	1000	(622)	1000	(622)	1000	(622)
Temperature	°C (°F)	130	(266)	130	(266)	120	(248)	140	(294)
SO_2 concentration	ppm**	1000		1000		1000		1000	
NO_x concentration	ppm	180		180		180		185	
Behind first stage adsorber									
SO_2 concentration	ppm	90		90		60		180	
NO_x concentration	ppm	130		130		138		130	
Behind second stage adsorber									
SO_2 concentration	ppm	Trace		Trace		Trace		Trace	
NO_x concentration	ppm	48		25		28		13	
NH_3 residual	ppm	6		17		15		50	
Behind ammonia adsorber									
NH_3 residual	ppm	Trace		Trace		Trace		Trace	
Total SO_2 removal efficiency %		>98		>98		>98		>98	
Total NO_x removal efficiency %		78		86		84		93	

Reference 5-Ebara

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*Water content in the flue gas is 10 percent

**SO_3 concentration is 10 ppm and total SO_3 removal efficiency is more than 98 percent.

TABLE 6. AVCO-EBARA PROCESS PROJECTED CAPITAL REQUIREMENTS

Unit Size (MW net)	2 x 500
Capacity Factor	65%
Coal Sulfur Content	4%
Pricing Level	December, 1982
SO_2 Emission Limit	1979 NSPS

Item	$/kW
Raw mateial receiving and storage	16.7
Flue gas treatment	80.0
Particulate removal credit	(18.3)
Reaction product separation and treatment	59.0
Flue gas supply	5.1
By-product storage	21.8
TOTAL process capital	164.3
General facilities @ 12.5% of 164.3	20.4
Engineering and home office fees @ 12.5% of 164.3	20.4
Project contingency @ 30% of 185	61.5
Process contingency @ 25% of 80.0 + 10% of 59.0	25.8
TOTAL plant cost	292.4
TOTAL plant investment	309.9
Royalty allowance @ 0.5% of 164.3	.84
Preproduction cost	13.0
Inventory and initial chemicals	9.6
Land	0
TOTAL capital requirement	333.34
Levelized fixed charges, mills/kWh:	9.4

TABLE 7. AVCO-EBARA PROCESS PROJECTED FIRST-YEAR OPERATING COSTS

Unit Size (MW net)	2 x 500
Capacity Factor	65%
Coal Sulfur Content	4%
Pricing Level	December, 1982
SO_2 Emission Limit	1979 NSPS

Item	Unit	Unit Rate ($)	Annual Quantity	Annual ($000)
Operating labor	Manhour	18.45	40,000	738
Maintenance labor				3,081
Maintenance material				4,621
Admin./support @ 30% of (738 + 3081)				1,146
TOTAL Fixed O&M cost				9,586
Utilities				
Steam	10^3 lb	3.02	160,000	484
Water	10^3 gal	.6	164,000	99
Electricity[a]	kWh	.042	223.5 x 10^6	9,466
Fuel oil or natural gas	10^6 Btu	8.22	76,000	626
Chemicals				
Anhydrous ammonia	Ton	262	124,000	32,559
Kaolin	Ton	133	23,000	3,061
Other consumables (fixed allowance)				1,452
By-product	Ton		462,000	-
Total variable O&M cost				47,747
Total first-year operating cost				57,333
Levelized operating cost, mills/kWh:		23.5		

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a) Annual quantity reflects a particulate removal credit of 735 kW.

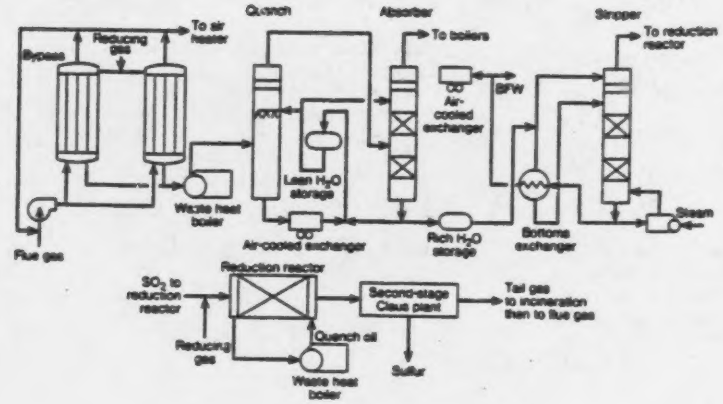

Quench
Absorber
Stripper
To air heater
Reducing gas
Bypass
To boilers
To reduction reactor
Air-cooled exchanger
BFW
Lean H_2O storage
Waste heat boiler
Flue gas
Air-cooled exchanger
Rich H_2O storage
Bottoms exchanger
Steam
Reduction reactor
SO_2 to reduction reactor
Second-stage Claus plant
Tail gas to incineration then to flue gas
Reducing gas
Quench oil
Waste heat boiler
Sulfur

Figure 1. Shell copper oxide process flow diagram. [1]

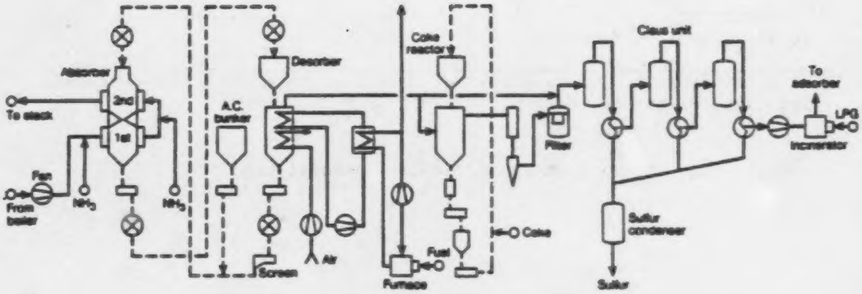

Absorber
2nd
1st
To stack
Fan
From boiler
NH_3
NH_3
A.C. bunker
Desorber
Screen
Air
Coke reactor
Furnace
Fuel
Coke
Filter
Claus unit
To adsorber
LPG
Incinerator
Sulfur condenser
Sulfur

Figure 2. Mitsui-BF dry desox denox process with sulfur recovery process. [5]

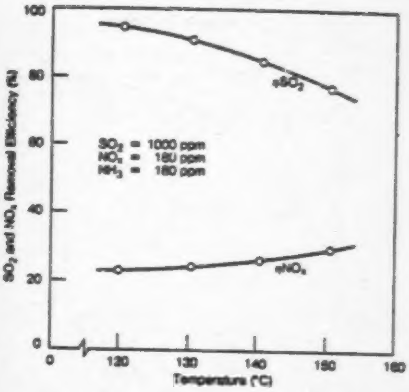


Figure 3. Effect of flue gas temperature on SO_2 and NO_x removal efficiency in the first-stage adsorber (Mitsui-BF).

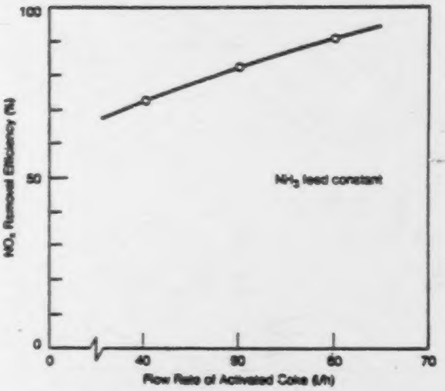


Figure 4. Relation between NO_x removal efficiency and flow rate of activated coke in the second stage adsorber Mitsui-BF.

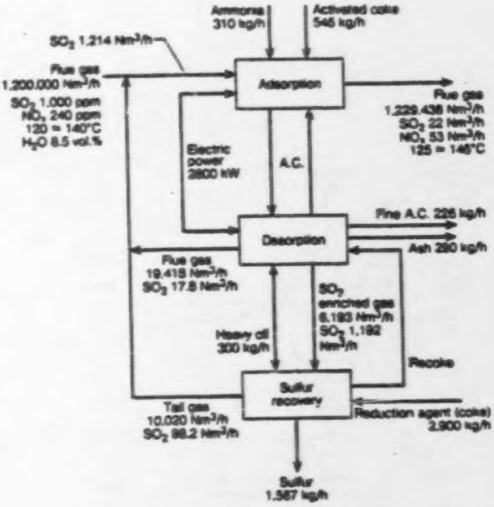

Ammonia 310 kg/h
Activated coke 545 kg/h
SO₂ 1,214 Nm³/h
Flue gas 1,200,000 Nm³/h SO₂ 1,000 ppm NOₓ 240 ppm 120 ~ 140°C H₂O 8.5 vol.%
Adsorption
Flue gas 1,229,438 Nm³/h SO₂ 22 Nm³/h NOₓ 53 Nm³/h 125 ~ 145°C
Electric power 2800 kW
A.C.
Desorption
Fine A.C. 225 kg/h
Ash 290 kg/h
Flue gas 19,415 Nm³/h SO₂ 17.8 Nm³/h
SO₂ enriched gas 8,193 Nm³/h SO₂ 1,192 Nm³/h
Heavy oil 300 kg/h
Recoke
Sulfur recovery
Tail gas 10,020 Nm³/h SO₂ 98.2 Nm³/h
Reduction agent (coke) 2,900 kg/h
Sulfur 1,587 kg/h

Figure 5. Material and energy balance dry desox denox system with sulfur recovery for a power plant of 1,200,000 Nm^3/h flue gas (Mitsui-BF).

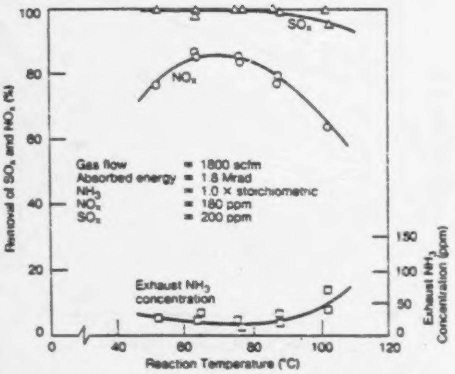


Figure 6. Removal of SO_x–NO_x and exhaust NH_3 concentration versus reaction temperature. [3]

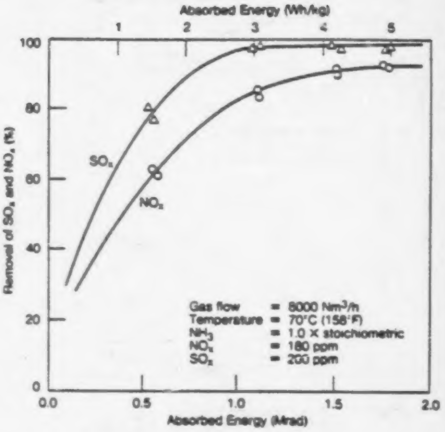


Figure 7. SO_x and NO_x removal versus electron dose Avco-Ebera. [2]

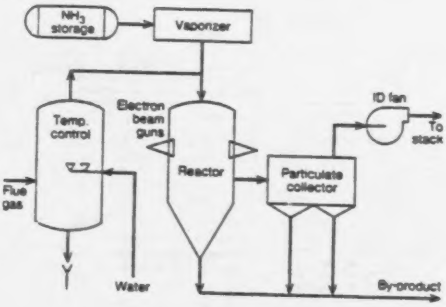

NH_3 storage
Vaporizer
Temp. control
Electron beam guns
ID fan
To stack
Reactor
Particulate collector
Flue gas
Water
By-product

Figure 8. E-beam SO_x/NO_x/particulate removal system schematic diagram. [6]

Taken from a Department of Commerce report on NOx abatement processes from stationary sources in Japan, this study describes wet processes that remove NOx only and SOx and NOx simultaneously. Important to notice is the variation in costs for individual systems. For example, a simultaneous SOx/NOx system (Ishikawajima-Harima Heavy Industries) seems to be more economical than combining SCR and FGD when burning high-sulfur oil. The same process (IHI) is not cost-effective, however, when burning coal, due to the amount of ozone required.

SECTION 7

WET PROCESSES FOR NO_x AND SIMULTANEOUS NO_x AND SO_x REMOVAL

7.1 GENERAL DESCRIPTION

7.1.1 Major Processes and Plants

Major commercial and test plants for NO_x and simultaneous NO_x and SO_x removal by wet processes are listed in Table 7-1. The plants are relatively small; the largest one has a capacity of 100,000 Nm^3/hr (33 MW equivalent).

The major wet processes may be classified as follows:

1) Oxidation-absorption

2) Absorption-oxidation

3) Oxidation-absorption-reduction (simultaneous removal of NO_x and SO_x)

4) Absorption-reduction (simultaneous removal of NO_x and SO_x)

Essentially all of the NO_x in the combustion gas is in the form of NO, which has poor reactivity and is not readily absorbed by most absorbents. NO is oxidized in NO_2 in air, but the oxidation occurs slowly. In processes (1) and (3) oxidizing agents are used to convert NO to NO_2 or N_2O_5 which are more readily absorbed. In the absorption oxidation process, NO is absorbed in a solution containing an oxidizing agent such as sodium

TABLE 7-1. MAJOR PLANTS FOR NO_x REMOVAL FROM FLUE GAS BY WET PROCESSES

Process Developer	Type of Process	Plant Owner	Plant Site	Capacity, Nm³/hr	Source of Gas	Completion	By-Product
Tokyo Electric-Mitsubishi H.I.	Oxidation absorption	Tokyo Electric	Minami-Yokohama	2,000[a]	Boiler[b]	Dec 1973	HNO_3
Tokyo Electric-Mitsubishi H.I.	Oxidation absorption	Tokyo Electric	Minami-Yokohama	100,000[a]	Boiler[b]	Oct 1974	HNO_3
Kawasaki H.I.	Oxidation absorption	EPDC	Takehara	5,000[a]	Boiler[c]	Dec 1975	Gypsum, $Ca(NO_3)_2$
Nissan Engineering	Absorption oxidation	Iijima Metal	Tokyo	2,000	Pickling	July 1973	$NaNO_2$
Nissan Engineering	Absorption oxidation	Nissan Chemical	Toyama	3,000	HNO_3 Plant	Mar 1975	$NaNO_3$, NO
Mitsubishi Metal-MKK-Nihon Chem.	Absorption oxidation	Mitsubishi Metal	Omiya	4,000[a]	Boiler[d]	Dec 1974	HNO_3
Kobe Steel	Absorption oxidation	Kobe Steel	Kakogawa	1,000[a]	Furnace[e]	Dec 1973	Gypsum, N_2
Kobe Steel	Absorption oxidation	Kobe Steel	Kakogawa	50,000[a]	Furnace[e]	Mar 1976	Gypsum, N_2
Hodogaya	Absorption oxidation	Hodogaya	Koriyama	14,000[a]	Furnace[e]	Oct 1975	$NaNO_2$, NaCl
Sumitomo Metal-Fujikasui	Oxidation reduction	Sumitomo Metal	Amagasaki	62,000	Boiler[d]	Dec 1973	$NaNO_3$, NaCl, Na_2SO_4
Sumitomo Metal-Fujikasui	Oxidation reduction	Toshin Steel	Fuji	110,000	Furnace[f]	Dec 1974	$NaNO_3$, NaCl, Na_2SO_4
Sumitomo Metal-Fujikasui	Oxidation reduction	Sumitomo Metal	Osaka	30,000	Boiler[d]	Dec 1974	$NaNO_3$, NaCl, Na_2SO_4
Sumitomo Metal-Fujikasui	Oxidation reduction	Sumitomo Metal	Kashima	25,000[a]	Sintering machine	1975	Gypsum, $CaCl_2$, Ca, NO
Osaka Soda	Oxidation reduction	Osaka Soda	Amagasaki	50,000	Boiler[d]	Mar 1974	$NaNO_3$, NaCl, Na_2SO_4
Shirogane	Oxidation reduction	Mitsui Sugar	Kawasaki	40,000	Boiler[d]	Aug 1974	Na_2SO_4, $NaNO_3$
Chiyoda	Oxidation reduction	Chiyoda	Kawasaki	1,000[a]	Boiler[d]	1973	Gypsum, $Ca(NO_3)_2$
Mitsubishi H.I.	Oxidation reduction	Mitsubishi H.I.	Hiroshima	2,000[a]	Boiler[d]	Dec 1974	Gypsum, NH_3
Ishikawajima H.I.	Oxidation reduction	Ishikawajima H.I.	Yokohama	5,000[a]	Boiler[d]	[illegible] 1975	Gypsum, N_2
Kureha Chemical	Absorption reduction	Kureha Chem.	Nishiki	5,000[a]	Boiler[d]	Apr 1975	Gypsum, N_2
Chisso Corp.	Absorption reduction	Chisso P.C.	Goi	300[a]	Boiler[d]	1976	$(NH_4)_2SO_4$
Mitsui Engineering	Absorption reduction	Mitsui P.C.	Chiba	150[a]	Boiler[d]	1974	H_2SO_4, N_2
Asahi Chemical	Absorption reduction	Asahi Chemical	Mizushima	500[a]	Boiler[d]	1974	Gypsum, N_2

[a] Test plants; others are commercial plants.
[b] Gas-fired boiler.
[c] Coal-fired boiler.
[d] Oil-fired boiler.
[e] Iron-ore sintering machine.
[f] Metal-heating furnace.

or potassium permanganate, while in the absorption reduction process NO is absorbed in a liquor containing catalysts, as will be described in Section 7.1.4.

Little progress has been observed with wet processes in the past year. Many of the pilot plants have discontinued operation for the following reasons: 1) the processes are not as simple as selective catalytic reduction, and 2) many of the processes by-produce nitrates or ammonia. The Japanese fertilizer industry has suffered from excessive ammonia production capacity. Moreover, Japanese farmers have used little nitrate fertilizer because the major crop is paddy rice for which nitrate fertilizers are not suitable.

Some of the wet processes, however, are capable of removing NO_x and SO_x simultaneously and converting NO_x to nitrogen or ammonia which can be used for fertilizer in many other countries, in contrast to the dry processes which consume a considerable amount of ammonia. Existing FGD plants can be modified in order to remove NO_x simultaneously by adding a few vessels. It is desirable that the wet processes will be improved to produce a simpler system and more favorable economics.

7.1.2 Oxidizing Agents

Ozone (O_3) and chlorine dioxide (ClO_2) are used mainly for the oxidation of NO in the gaseous phase. They oxidize NO to NO_2 within a second but barely oxidize SO_2 to SO_3. Ozone can oxidize NO to N_2O_5 when added in excess.

$$NO + O_3 \longrightarrow NO_2 + O_2 \qquad (7\text{-}1)$$

$$2NO + 3O_3 \longrightarrow N_2O_5 + 3O_2 \qquad (7\text{-}2)$$

Ozone is fairly expensive; it costs 330 to 400 yen per kilogram and about 0.4 yen/kWh ($1.65 to $2.00/kg, $0.002/kWh) to treat a flue gas from a boiler containing 150 ppm of NO. A large scale ozone generator with a capacity of 100 kilograms per hour of ozone is near completion. It can treat about 230,000 Nm^3/hr of flue gas (76 MW equivalent) containing 200 ppm NO. The cost of ozone is expected to decrease to some extent with the large generator. The cost of chlorine dioxide is 30 to 40 percent lower than that of ozone, but chlorine dioxide introduces hydrochloric and nitric acids, which complicate the system.

$$2NO + ClO_2 + H_2O \longrightarrow NO_2 + HNO_3 + HCl \qquad (7\text{-}3)$$

Solutions of potassium and sodium permanganates, sodium and calcium hypochlorites, and hydrogen peroxide have been used for oxidation in the liquid phase, but these chemicals are also expensive.

7.1.3 Oxidation-Absorption and Absorption-Oxidation Processes

In oxidation-absorption processes the NO is first oxidized with a gaseous oxidizing agent and then absorbed. In absorption-oxidation processes the NO is absorbed in a solution containing an oxidizing agent. Usually NO_x absorption occurs more slowly in the latter case because NO must be absorbed in the liquor before it can be oxidized.

Most plants which use nitric acid for such processes as metal washing emit a gas fairly rich in NO_x (1000 to 10,000 ppm). However, the amount of gas is not great (500 to 5000 Nm^3/hr). In many of the plants, all or part of the NO is oxidized to NO_2, and the gas is absorbed in a sodium hydroxide solution. Activated carbon is used in some of the plants as a catalyst for NO oxidation by air. In other plants NO_x is absorbed in a solution containing an oxidizing agent such as NaClO or H_2O_2. In both cases the resulting liquor, containing nitrate and nitrite, is sent to a wastewater treatment system. Some plants use activated sludge treatment to reduce the nitrogen content of the water. Such processes cannot be applied on a large scale because the treatment does not sufficiently remove nitrogen compounds from the wastewater.

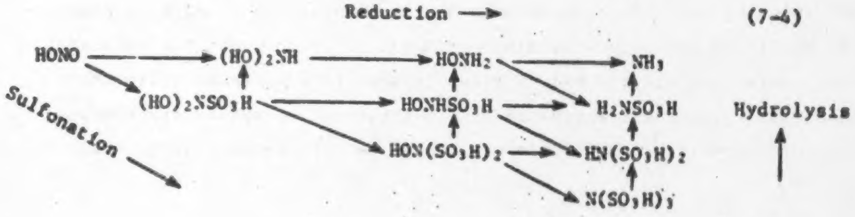
Tests were made in pilot plants to recover nitric acid for industrial use (see Section 7.2.2) or to recover potassium or calcium nitrate for fertilizer (see Sections 7.9.1 and 7.9.2). Those processes do not seem promising because of the high cost and the limited demand for the by-products.

7.1.4 Oxidation-Absorption-Reduction and Absorption-Reduction Processes (Simultaneous Removal)

Since 1973 many oxidation-absorption-reduction and absorption-reduction processes have been developed in which NO_x and SO_x are absorbed simultaneously. In the oxidation-absorption-reduction process, NO is first oxidized and then absorbed together with SO_x by lime/limestone scrubbing. In the absorption-reduction process NO is absorbed with SO_x in a liquor containing ferrous ion, which can form an adduct with NO. Usually EDTA (ethylenediamine tetraacetic acid, a chelating compound whose present cost in Japan is about 800,000 yen/t, $4000/t) is added to promote the reaction of ferrous ion with NO (Figures 7-1 and 7-2).[1]

In both cases, various reactions occur in the liquor or slurry and result in the reduction of NO_x by SO_2 (or sulfite) to NH_3 through imidodisulfonic acid ($HN(SO_3H)_2$), sulfamic acid (H_2NSO_3H), or a salt of either acid. The reaction sequence may be expressed in the following way.[2]

(7-4)



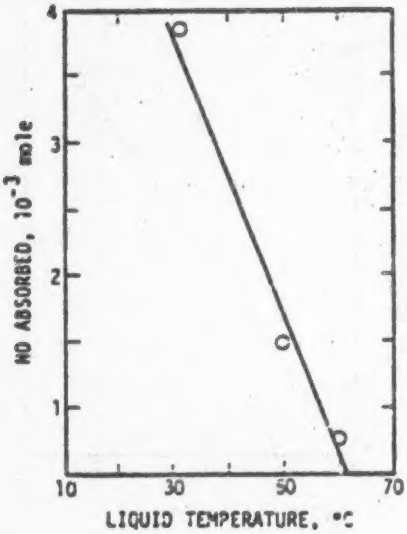


Figure 7-1. NO absorption in EDTA-Fe(II)liquor (0.01 mole/liter).*

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* Amount of liquor, 1 liter; gas flow, 1.5 liter/min; pH of liquor, 5.65; initial NO concentration, 275 ppm.

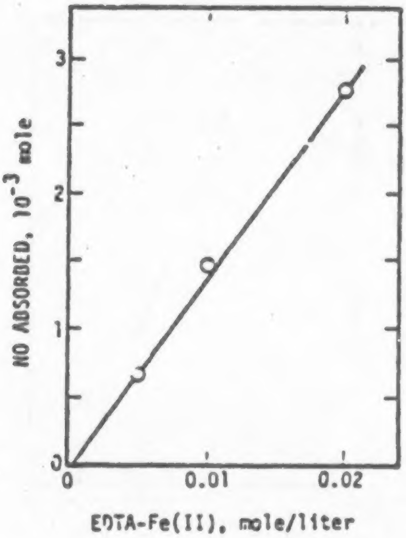

Figure 7-2. EDTA-Fe(II) concentration and NO absorption at 50°C.*

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* Amount of liquor, 1 liter; gas flow, 1.5 liter/min; pH of liquor, 5.65; initial NO concentration, 275 ppm.

Although the chemistry is complex, the overall reactions may be expressed simply in the following equations:

$$2NO_2 + 7SO_2 + 7CaCO_3 + 3H_2O = 7CaSO_4 + 2NH_3 + 7CO_2 \qquad (7\text{-}5)$$

([illegible]ation reduction process, limestone scrubbing)

$$2NO + 5SO_2 + 8SO_2 + 3H_2O = 5(NH_4)_2SO_4 \qquad (7\text{-}6)$$

(absorption reduction process, ammonia scrubbing)

The reactions indicate that 3.5 moles of SO_2 is needed for 1 mole of NO_2 and 2.5 moles of SO_2 is needed for 1 mole of NO. Since flue gas contains O_2 which oxidizes a considerable portion of SO_2 in a scrubber liquor, 4-5 moles of SO_2 is actually needed to attain 80 percent NO_x removal.

A portion of the NO_x can be reduced to N_2 in an aqueous solution. The reactions are complex but may be simply described as:

$$NH_2SO_3H + NO_2^- \longrightarrow N_2 + HSO_4^- + H_2O \qquad (7\text{-}7)$$

Tanaka has found that a compound $Na_2SO_3 \cdot 2NO$ is formed when NO is absorbed in an $NaSO_3$ solution. The compound is stable at a pH above 8, but it decomposes to form Na_2SO_4 and N_2O at lower pH. It is possible that in addition to N_2 or NH_3, N_2O also is formed in some of the wet processes.

$$Na_2SO_3 \cdot 2NO \longrightarrow Na_2SO_4 + N_2O \qquad (7\text{-}8)$$

The presence of EDTA and ferrous ion prevents the formation of N_2O.

In some of the processes a considerable portion of the NO_x remains in the resulting liquor as nitrite and nitrate, which would cause a problem in wastewater treatment.

The advantage of such wet processes over dry processes is that they can simultaneously remove SO_2 and NO_x without the problems of particulates and ammonium bisulfate. They have not yet been commercialized on a large scale. Five relatively small commercial plants are in operation (Table 7-1).

7.2 MHI WET PROCESSES[3]

7.2.1 Introduction

Mitsubishi Heavy Industries (MHI), the leading FGD process developer and FGD plant constructor, has developed two wet processes for NO_x and simultaneous SO_x and NO_x removal, both using ozone as the oxidizing agent. The NO_x removal process uses water absorption to produce nitric acid; it was tested in cooperation with Tokyo Electric Power at Minamiyokohama Power Station. The simultaneous removal process uses lime/limestone scrubbing; it was tested at MHI's Hiroshima Research Institute.

7.2.2 Nitric Acid By-Production Process

A flowsheet of the process is shown in Figure 7-3. The flue gas (100,000 Nm^3/hr) from an LNG-fired boiler containing about 100 ppm NO_x is cooled with water spray, injected with ozone to oxidize NO to N_2O_5, and then washed with water in an absorber. More than 90 percent of the NO_x is removed when more than 1.7 moles of ozone are used for every mole of NO. An approximately 10 percent nitric acid solution is obtained which is concentrated to 60 percent for industrial use. The gas discharged from the absorber contains a small amount of ozone, which is removed by treatment with calcium sulfite slurry containing a small amount of catalyst. Major facilities of the plant are shown in Table 7-2.

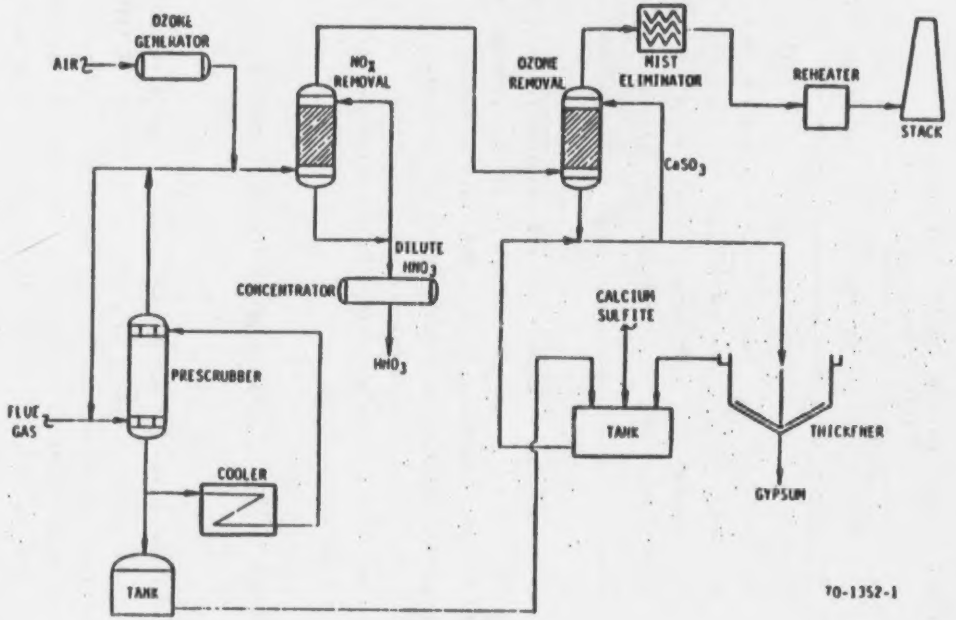
OZONE GENERATOR
AIR
FLUE GAS
PRESCRUBBER
COOLER
TANK
NO_x REMOVAL
DILUTE HNO_3
CONCENTRATOR
HNO_3
OZONE REMOVAL
MIST ELIMINATOR
REHEATER
STACK
$CaSO_3$
CALCIUM SULFITE
TANK
THICKENER
GYPSUM
70-1352-1

Figure 7-3. Flowsheet of Tokyo Electric-MHI oxidation absorption process.

TABLE 7-2. SPECIFICATIONS FOR THE TEST PLANT
(TOKYO ELECTRIC-MHI PROCESS)

	Material	Structure	Dimensions (m)
Prescrubber	SUS 304L	Packed Tower	4 x 4 x 11
NO_x Absorber	SUS 304L	Packed Tower	4 x 4 x 19.5
Ozone Absorber	SUS 304L	Packed Tower	4 x 4 x 19.5
Mist Eliminator	Polypropylene	Chevron	2.6 x 1.8 x 3
Blower	SS 60 (Impeller)	Centrifugal	370 kW

Operation parameters are listed in Table 7-3 and compared to those for the simultaneous removal plant. The nitric acid by-production process plant was operated for about 2 years beginning in late 1974 with virtually no problem. Ozone leaving the absorber in a small amount was entirely caught by the sulfite slurry and gypsum of good quality was by-produced. The process, however, is more costly than SCR which can be easily used for clean gas. The plant operation has been shut down.

7.2.3 Simultaneous Removal Process (Oxidation-Absorption-Reduction Process)

Another wet simultaneous removal process is a modification of the wet lime/limestone process as shown in Figure 7-4. Flue gas from an oil-fired boiler (2000 Nm^3/hr) is first cooled to 60°C by a prescrubber. Ozone is introduced into the flue gas prior to scrubbing. A water-soluble catalyst in a lime/limestone slurry is added to promote the oxidation of NO. More than 85 percent of the NO_x and more than 95 percent of the SO_2 are removed when more than 4 moles of SO_2 per mole of NO_x are present in the gas. The slurry discharged from the scrubber contains gypsum and a small amount of unreacted lime or limestone and soluble nitrogen-sulfur (N-S) compounds (reactions 7-9 - 7-11). The unreacted lime or limestone is treated with

TABLE 7-3. PILOT PLANTS BY MHI WET PROCESSES

	Nitric Acid By-Production	Simultaneous Removal
Gas Source	LNG-Fired Boiler	Oil-Fired Boiler
Capacity (Nm^3/hr)	100,000	2000
O_3/NO_x Mole Ratio	1.7-2.0	0.9-1.0
NO_x (inlet, ppm)	100	150-250
NO_x Removal (%)	Over 90	75-92
SO_2 (inlet, ppm)	None	500-1200
SO_2 removal (%)	--	90-99
Particulates (inlet, mg/Nm^3)	Trace	90-120
Prescrubber Type	Packed Tower	Spray Tower
Absorber Type	Packed Tower	Packed Tower
pH of Scrubber Liquor	Below 1	4-6
Ozone Remover Type	Packed Tower	--
By-Product	HNO_3, Gypsum	Gypsum
Year Completed	1974	1975

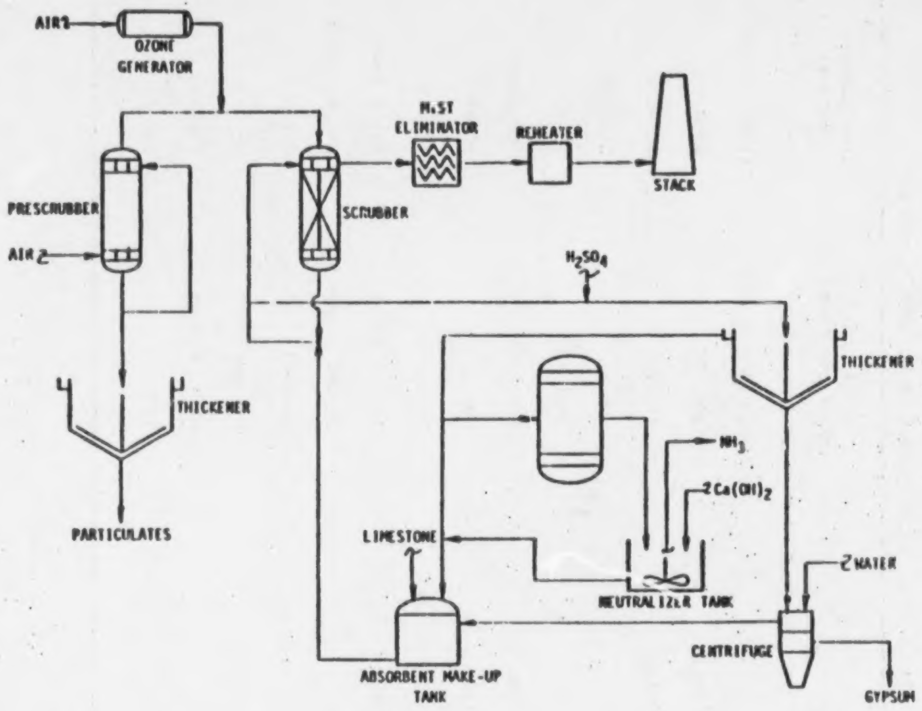
AIR
OZONE GENERATOR
PRESCRUBBER
AIR
THICKENER
PARTICULATES
SCRUBBER
MIST ELIMINATOR
REHEATER
STACK
H_2SO_4
THICKENER
NH_3
$Ca(OH)_2$
LIMESTONE
NEUTRALIZER TANK
WATER
ABSORBENT MAKE-UP TANK
CENTRIFUGE
GYPSUM

Figure 7-4. MHI simultaneous removal process.

sulfuric acid to increase the purity of by-product gypsum; the slurry is centrifuged to recover gypsum. Most of the liquor is returned to the absorber after lime or limestone is added; a small portion of the liquor is treated in a reactor at 100-130°C to decompose more than 95 percent of the N-S compounds (reactions 7-12 - 7-15). A small portion of the nitrogen is converted to N_2; most of it is converted to NH_4HSO_4 which is then treated with lime to recover gypsum and gaseous ammonia. The gypsum and ammonia, 5-10 percent concentration, may be recovered and used (reaction 7-16).

$$2NO_2 + Ca(OH)_2 + CaSO_3 \cdot \tfrac{1}{2} H_2O + \tfrac{1}{2} H_2O \longrightarrow Ca(NO_2)_2 + CaSO_4 \cdot 2H_2O \qquad (7\text{-}9)$$

$$Ca(NO_2)_2 + 4(CaSO_3 \cdot \tfrac{1}{2}H_2O) + 2H_2O \longrightarrow 2CaNOH(SO_3)_2 + 3Ca(OH)_2 \qquad (7\text{-}10)$$

$$CaNOH(SO_3)_2 + CaSO_3 \cdot \tfrac{1}{2}H_2O + \tfrac{3}{2} H_2O \longrightarrow CaNH(SO_3)_2 + CaSO_4 \cdot 2H_2O \qquad (7\text{-}11)$$

$$2Ca[NOH(SO_3)_2] + 2H_2O \longrightarrow Ca(NOH \cdot HSO_3)_2 + Ca(HSO_4)_2 \qquad (7\text{-}12)$$

$$Ca(NOH \cdot HSO_3)_2 + 2H_2O \longrightarrow \tfrac{2}{3} N_2 + CaSO_4 \cdot 2H_2O + \tfrac{2}{3} NH_4HSO_4 + \tfrac{1}{3}H_2SO_4 \qquad (7\text{-}13)$$

$$2Ca[NH(SO_3)_2] + 2H_2O \longrightarrow Ca(NH_2SO_3)_2 + Ca(HSO_4)_2 \qquad (7\text{-}14)$$

$$Ca(NH_2SO_3)_2 + Ca(HSO_4)_2 + 4H_2O \longrightarrow 2NH_4HSO_4 + 2CaSO_4 \cdot 2H_2O \qquad (7\text{-}15)$$

$$NH_4HSO_4 + Ca(OH)_2 \longrightarrow NH_3 + CaSO_4 \cdot 2H_2O \qquad (7\text{-}16)$$

Some of the operation data of the pilot plant are listed in Table 7-3. Specifications of the plant facilities are shown in Table 7-4.

TABLE 7-4. SPECIFICATIONS OF 2000 Nm^3/HR TEST PLANT FOR SIMULTANEOUS REMOVAL OF SO_x AND NO_x BY OXIDATION-ABSORPTION-REDUCTION

Prescrubber	Spray Tower	0.7 m diameter, 7 m height
Absorber	Grid Packed Tower	0.5 m diameter, 13 m height
Ozonizer	Mitsubishi Electric (OS-1200)	1200 g/hr*
Flue Gas Fan	Capacity 5000 Nm^3/hr	Pressure, 600 mm H_2O
Centrifuge	Basket Type	300 l/hr

* About half of the capacity is sufficient for the pilot plant.

The plant has been operated without appreciable problem. Nearly 1 mole of O_3 has been used for each mole of NO to obtain 85-90 percent NO_x removal efficiency. It is possible to obtain over 90 percent efficiency by using a larger amount of O_3. In this case, however, a small amount of nitrate is formed in the liquor, which makes wastewater treatment difficult. Unreacted ozone is entirely removed in the scrubber by the calcium sulfite slurry. There is no leakage of O_3 from the system.

Because of the oxidizing effect of NO_2 and O_3, virtually all of the sulfite is oxidized to gypsum in the scrubber. An oxidizing tower, which is usually installed for conventional lime/limestone gypsum processes, is not needed when the SO_2 content of flue gas is below 1200 ppm. For flue gas richer in SO_2, an oxidizer may be needed but should be substantially smaller than those in conventional FGD plants.

Gypsum derived from the sulfuric acid is less than 5 percent of the total by-product gypsum. Sulfuric acid is not needed when a high purity of gypsum is not required.

Both capital and operation costs for a commercial simultaneous removal process plant to treat flue gas from an oil-fired boiler containing about 150 ppm of NO_x are estimated by MHI to be about 40 percent higher than those for an SO_2 removal process plant. About 80 percent of the increased costs are due to the ozone generator.

7.2.4 Evaluation

A desirable feature of the by-product nitric acid process is that NO_x is converted to a useful by-product. The process, however, seems fairly costly and requires a large amount of ozone and many vessels. Moreover, it may not be applied to SO_x-containing gas. For NO_x removal from clean gas, SCR may be easier and less costly.

The simultaneous removal process is a fine process in which both NO_x and SO_2 are removed by the simple addition of ozone to the gas and a catalyst to the scrubber liquor. It is easy to modify the existing lime/limestone FGD plant for simultaneous removal.

A large disadvantage of the process is the high cost of ozone. The process may be useful for flue gas with a relatively low NO_x concentration such as that from an oil-fired boiler, but may be too expensive for an NO_x-rich gas such as that from a coal-fired boiler. Recovery of the by-product ammonia should be considered. The ammonia may be injected into the boiler to reduce NO_x. The simultaneous removal process may be applied to flue gas from a coal-fired boiler if NO_x concentration is reduced to 250 ppm by combustion control and to below 150 ppm by injecting the by-product ammonia into the boiler.

As in the case of conventional FGD processes, some wastewater would have to be purged to prevent the accumulation of impurities from the gas and lime or limestone. The recovery of catalyst from the wastewater may present a problem.

7.3 IHI OXIDATION-ABSORPTION-REDUCTION PROCESS (SIMULTANEOUS REMOVAL)

7.3.1 Introduction

Ishikawajima-Harima Heavy Industries (IHI), one of the largest producers of boilers, ships, and machines, has developed a wet simultaneous removal process with ozone oxidation and lime/limestone scrubbing to by-produce gypsum which converts NO_x to N_2. A pilot plant which can treat 5000 Nm^3/hr flue gas from an oil-fired boiler was operated in 1976 and 1977.

7.3.2 Process Description

A flowsheet of the IHI process is shown in Figure 7-5. Flue gas is cooled, injected with ozone to convert NO to NO_2, and treated in a scrubber with a lime/limestone slurry at pH 5-6 containing small amounts of $CuCl_2$ and NaCl as catalysts. Over 80 percent of the NO_x and over 95 percent of the SO_2 are absorbed at an L/G of over 8 liters/Nm^3. Most of the absorbed NO_x forms calcium sulfamate, a soluble compound (reaction 7-20), while a small portion of NO_2 is converted into various forms (reactions 7-21 - 7-24).

$$SO_2 + Ca(OH)_2 \longrightarrow CaSO_3 + H_2O \qquad (7\text{-}17)$$

$$SO_2 + CaCO_3 \longrightarrow CaSO_3 + CO_2 \qquad (7\text{-}18)$$

$$SO_2 + CaSO_3 + H_2O \longrightarrow Ca(HSO_3) \qquad (7\text{-}19)$$

$$2NO_2 + 5CaSO_3 + Ca(HSO_3) + 11H_2O \longrightarrow Ca(NH_2SO_3)_2 + 5(CaSO_4 \cdot 2H_2O) \qquad (7\text{-}20)$$

$$2NO_2 + 4CaSO_3 + 8H_2O \longrightarrow N_2 + 4(CaSO_4 \cdot 2H_2O) \qquad (7\text{-}21)$$

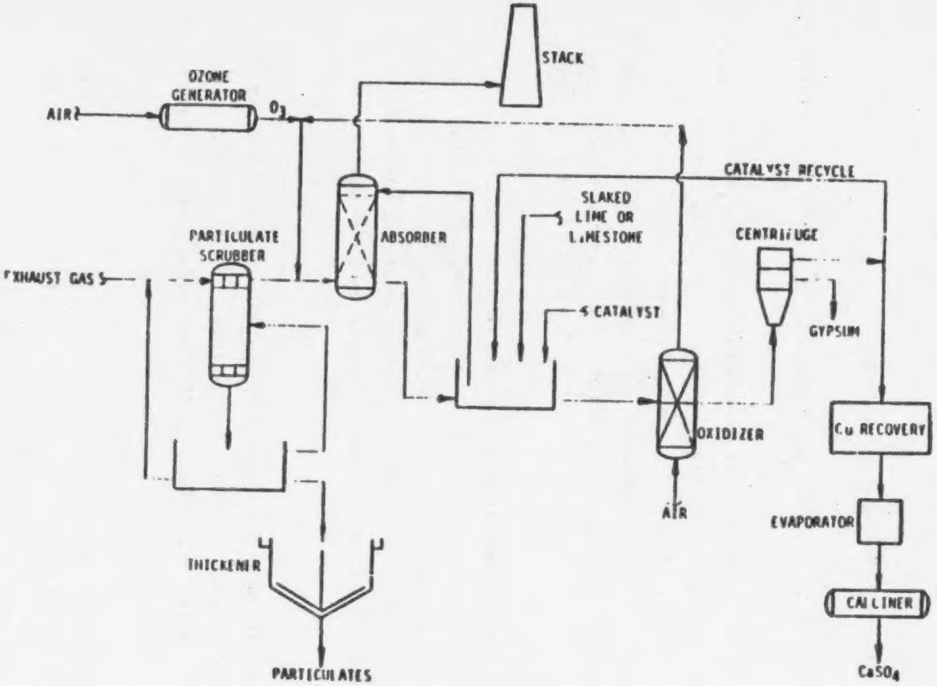
AIR
OZONE GENERATOR
O_3
STACK
EXHAUST GAS
PARTICULATE SCRUBBER
ABSORBER
SLAKED LINE OR LIMESTONE
CATALYST
CATALYST RECYCLE
CENTRIFUGE
GYPSIM
OXIDIZER
AIR
Cu RECOVERY
EVAPORATOR
CALCINER
$CaSO_4$
THICKENER
PARTICULATES

Figure 7-5. Flowsheet of the IHI process.

$$4NO_2 + 4CaSO_3 + 2H_2O \longrightarrow Ca(NO_2)_2 + Ca(NO_3)_2 + 2Ca(HSO_3)_2 \qquad (7\text{-}22)$$

$$2NO_2 + 3CaSO_3 + 6H_2O \longrightarrow N_2O + 3(CaSO_4 \cdot 2H_2O) \qquad (7\text{-}23)$$

$$Ca(NO_2)_2 + Ca(NH_2SO_3)_2 + 2H_2O \longrightarrow 2N_2 + 2(CaSO_4 \cdot 2H_2O) \qquad (7\text{-}24)$$

The relation between the catalyst concentration and NO_x removal efficiency is shown in Figure 7-6. The combination of $CuCl_2$ and NaCl is effective for NO_x removal. Figure 7-7 shows the effect of pH on SO_x and NO_x removal with and without the catalyst. A high pH increases SO_x removal but lowers NO_x removal. A pH between 5 and 6 is optimum for simultaneous removal.

The slurry leaving the absorber is treated with air bubbles if necessary to complete the oxidation of $CaSO_3$ to gypsum. The gypsum slurry is filtered to obtain a good quality by-product gypsum. Most of the liquor discharged from the centrifuge is returned to the absorber; the rest is sent to a thermal decomposition step after separation of copper by lime addition.

For the thermal decomposition, the liquor is first evaporated and dried to crystallize calcium sulfamate and nitrate. These are then calcined with limestone in a rotary kiln at about 900°C to produce calcium sulfate and nitrogen gas (reactions 7-25 and 7-26).

$$2Ca(NH_2SO_3)_2 + 4CaCO_3 + 2Ca(HSO_4)_2 + 3O_2 + 2H_2O \longrightarrow \qquad (7\text{-}25)$$
$$2N_2 + 8CaSO_4 + 4CO_2 + 8H_2O$$

$$Ca(NO_3)_2 + Ca(NH_2SO_3)_2 \longrightarrow 2N_2 + 2CaSO_4 + O_2 + 2H_2O \qquad (7\text{-}26)$$

The pilot plant was operated without appreciable problem. IHI also developed a process to hydrolize the sulfamate to by-produce ammonia. In

CONDITIONS

CONCENTRATION OF $CaSO_4$: 5 wt%
pH OF LIQUOR : 5.5
$NO_2/(NO + NO_2)$: 0.95

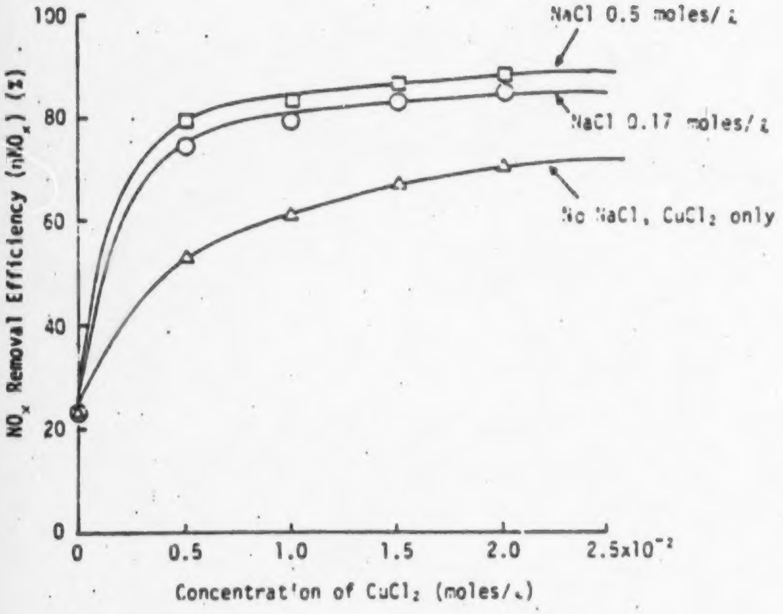

Figure 7-6. Effect of $CaCl_2$ and NaCl concentration on NO_x removal efficiency in the IHI process.

CONDITIONS

CONCENTRATION OF $CaSO_3$: 5 wt%
CONCENTRATION OF $CuCl_2$: 0.01 mole/ℓ
CONCENTRATION OF NaCl : 1 wt%
$NO_2/(NO + NO_2)$: 0.95

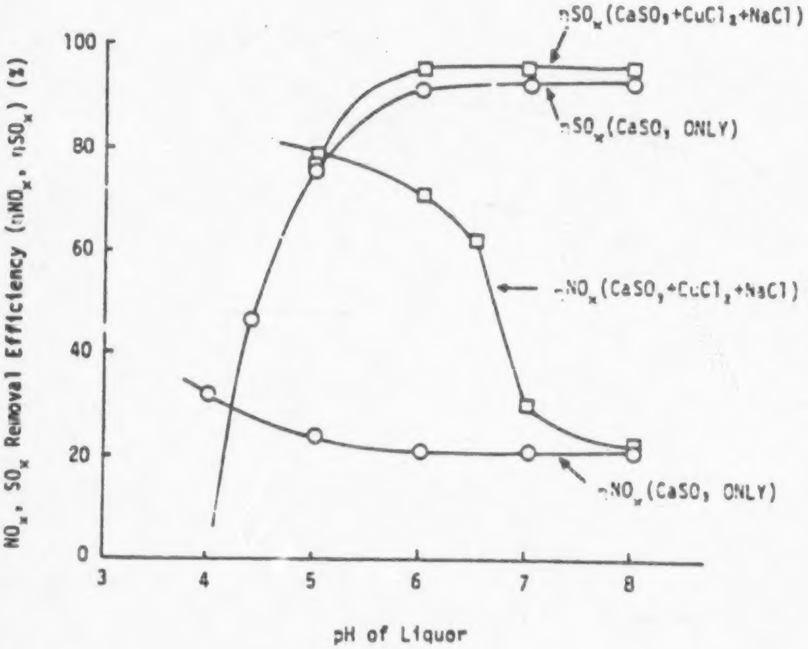

Figure 7-7. Effect of pH on SO_x and NO_x removal efficiency in the IHI process.

this case, calcium nitrate accumulates in the scrubber liquor and has to be separated.

7.3.3 Economics

Figure 7-8 shows investment cost and operating cost for the simultaneous removal of SO_x and NO_x for 8000 hours of annual operation. The figures include 7 years' depreciation. Roughly speaking, both investment and operation costs are about 40 percent more than those of FGD only. The figure also shows the incremental operating cost for NO_x removal (the difference between simultaneous removal and FGD). The operation costs may be considerably higher for NO_x-rich flue gas from a coal-fired boiler because a much larger amount of ozone may be needed.

7.3.4 Evaluation

IHI made extensive fundamental studies of the IHI process. The pilot plant was operated without appreciable problem and proved the reliability of the process. However, there is no plan yet to install a commercial plant based on the process. The major reason for this hesitance is the current policy of power companies to use LNG or low-sulfur oil for new boilers with selective catalytic reduction (SCR) of NO_x without FGD.

For flue gas from high-sulfur oil, the IHI process might be more economical than a combination of SCR and wet FGD. The process may not be useful for flue gas from a coal-fired boiler because the large amount of ozone required would make the process too costly.

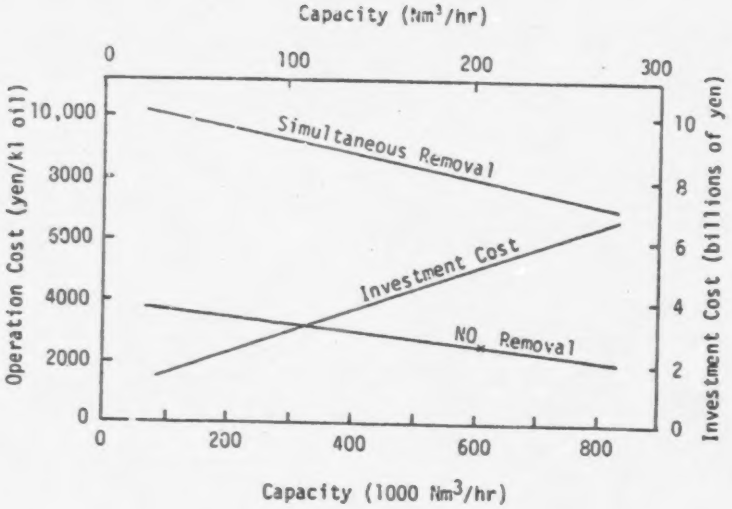

Figure 7-8. Investment and operation costs.*

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* Inlet SO_x, 1500 ppm; inlet NO_x, 180 ppm; yearly operation, 8000 hours.

The Environmental Protection Agency (EPA) is pursuing development of LIMB technology, as a promising way to reduce both NOx and SOx at very low costs. This EPA report outlines the systems and describes its potential market uses and future development needs. Discussion of the development to date is described, including activities in West Germany. The EPA is trying to guide the development of LIMB to maximize its potential application.

LIMB - EPA'S PROGRAM TO DEVELOP LOW-COST SO_x/NO_x CONTROL

I. INTRODUCTION

The LIMB Program - What is it?

The Limestone Injection/Multistaged Burners (LIMB) program is an EPA-sponsored effort to develop an effective but inexpensive emission control technology for pulverized coal boilers that will simultaneously remove sulfur oxides (SO_x) and nitrogen oxides (NO_x) from boiler flue gases. The technology is based on the use of low NO_x combustion techniques in combination with dry limestone injection into the furnace (combustion chamber) for simultaneous SO_x control.

The program goal is to develop the technology to obtain substantial reductions in SO_x and NO_x emissions for a capital investment cost of 30-40 $/kilowatt - about 1/5 of the cost of conventional flue gas desulfurization (FGD) systems. For cases involving retrofit of the technology on existing pulverized coal boilers, the goal is to achieve 50 to 70 percent SO_x control and NO_x emission levels of 130-170 ng/J (0.3-0.4 lb/million Btu) thermal input. For new boiler applications the goal is to achieve even higher SO_x removals while achieving NO_x emission levels of 86-130 ng/J (0.2-0.3 lb/million Btu). In addition, for new systems there exist the possibility of achieving synergism between the limestone injection process and various tailend cleanup devices for both particulate and SO_x thus producing a hybrid system which will be low cost and very effective (state-of-the-art) for combined SO_x, NO_x, and particulate. Such a process might combine, for instance, low NO_x burners, limestone injection, and an SO_x-particulate control system such as the spray dryer-baghouse combination.

II. Background

The concept of sorbent injection into furnaces for SO_x control is not a new concept. In the early development (1965-1970) of SO_x control technology, it was one of the alternatives considered for achieving high removal efficiency

(90%) at low cost. Extensive full scale testing was carried out on a number of boiler types with limited success. Typically, SO_x removals ranged from 20 to 50% for reasonable sorbent loadings (i.e. about 2 times stoichiometric calcium to sulfur), which was far short of the goal. The reasons most commonly quoted for not achieving the goal were: 1) high temperature deadburning of sorbent and decomposition of the sorbent sulfur complex; and 2) inefficient mixing of sorbent with combustion gases.

Interest in the technology was revived as a result of extensive development work on advanced design low NO_x coal burners sponsored by the EPA and others. To achieve low NO_x emissions, the mixing of coal and air are delayed which results in reduced peak temperatures and altered chemical conditions in the reaction zone. As a result, in 1978 a short series of trials was carried out at Energy and Environmental Research Corporation under EPA sponsorship to study sorbent injection through an advanced low NO_x burner. SO_x removals of up to 70% were achieved with limestone injection at a calcium to sulfur stoichiometry of two. These results were for an experimental 6 x 10^6 Btu/hr burner operating in a small experimental tunnel at conditions designed to minimize NO_x, and no optimization of SO_x capture was attempted.

III. Incentive for LIMB Program

Based on the new pilot scale work with sorbent injection through the low NO_x burner, some hypotheses were formulated to explain the differences from previous results. These revolve around the altered combustion conditions generated by the staged combustion burner:

1) The delayed mixing tends to spread out the heat release and thereby lower both peak and average temperatures in the reaction zone;

2) The fuel rich core of the burner may provide an environment where sulfur is present as reduced species (H_2S and COS). Based on kinetics, thermodynamic equilibrium and steric factors, capture as the sulfide appears beneficial. At higher temperatures sulfide species are more stable than sulfate and also have increased rates of formation;

3) Intimate contact of sorbent and sulfur in the burner zone may improve capture;

4) Activation of the sorbent in the reaction zone may give better capture.

While the limited pilot scale work was encouraging, additional work was clearly necessary to define potential in practical systems. The limited capture relative to flue gas desulfurization systems did not provide strong incentive for stand alone application to new boilers; however, the potential for combined systems appeared attractive.

A strong incentive for further development of the approach emerged from an assessment of strategies available for retrofit control of acid rain precursors. SO_x and NO_x from pulverized coal fired steam generators were major precursors to be considered. Based on a preliminary cost analysis it appeared that a LIMB technology to give 70% NO_x control and 50% SO_x control at a sorbent to sulfur ratio of two would be highly beneficial, especially when cost was considered. On that basis in late 1979 EPA began laying the ground work of a development program for LIMB.

IV. Program Identification

To define a LIMB program, it is necessary to examine the elements of the problem specific to the U.S. The key elements are as follows:

1) The U.S. boiler population is made up to two main classes of boilers: those with rapid mixing burners (pre NSPS wall fired boilers); and those with delayed mixing burners (Tangential, arch fired and post NSPS wall fired). Within each class there are a number of additional design variables which eust be considered. To satisfy retrofit applications, the program must provide the basis for application to the full spectrum of boilers.

2) The U.S. has a wide spectrum of coals ranging from low heating value lignites to high heating value bituminous. The properties of the coals vary widely, especially the sulfur level, distribution of

sulfur type, and alkali metal content. The technology must be assessed for the full spectrum of coals, although emphasis should be placed on high sulfur eastern bituminous.

3) System optimization must be considered for new applications. High efficiency SO_x control is required to meet current NSPS (for new boilers), which will require either an optimization of the boiler/burner design or the use of a post capture system for additional SO_x control. The program must provide a basis for new system design.

To satisfy these key elements, a program was structured to encompass five areas as follows:

1) Basic understanding of the reaction mechanisms is essential to generalization of the technology. While a great deal of work has been done on the calcium-SO_2 reactions, the available data on the sulfide reaction mechanism is insufficient. In addition very little work has been done on sorbent sulfur reactions under actual coal combustion conditions. This information will provide input both to the optimization of conventional systems and for innovative approaches for new systems.

2) Systematic small bench and pilot scale development of combustion technology is necessary for the wide range of boiler and fuel types. There are a wide range of variables which must be examined, including: burner mixing, sorbent injection; thermal environment; sorbent type and stoichiometry; coal properties; and others. To quantify the effects, experiments must be run under well controlled conditions in as economical a manner as possible. The results must address not only SO_x and NO_x control, but also particulate related system effects such as slagging, fouling, and collection device efficiency.

3) Large pilot scale testing is required to extrapolate the performance of small scale experiments to practical scale. The

testing should include optimization of full scale (nominal 100 x 10^6 Btu/hr) commercially-available and advanced second generation low NO_x coal burners in an experimental system simulating practical boilers. These tests should include SO_x optimization as well as an evaluation of NO_x/SO_x tradeoff. Emphasis will be placed on burners suitable for retrofit applications. This portion of the program will be carried out with direct cooperation and participation of boiler and/or burner manufacturers. It will provide the final confirmation of the potential of LIMB technology for field application.

4) Detailed process analysis is required to extrapolate the development data to practical boiler application and to identify potential operability/reliability problems for typical systems. It should also provide reliable cost estimates for both new and retrofit application.

5) The final step in the process must be the field application of the technology to representative bgilers. The LIMB program does not encompass field evaluation of sorbent injection technology; however, it will provide design guidance for efforts by private industry. Ongoing field evaluation of low NO_x burners on wall- and tangentially-fired boilers will be completed.

V. LIMB program

To implement the approach discussed above, a LIMB program has been defined to address the key issues. The following provides a brief description of the project activities underway or planned for each area.

A. Basic understanding of chemical kinetics of fuel rich sulfide chemistry will be provided by a combination of in-house research and cooperative agreements. Specific elements are as follows:

1) Evaluation of reaction rates of reduced sulfur species (H_2S & COS) with a range of sorbents under well controlled conditions will be

conducted in fixed bed and flow reactors. The effects of combustion will also be assessed.

2) The role of sorbent/coal contacting will be assessed by laboratory scale experiments where the active alkali metal is associated with the coal.

B. Bench and small pilot-scale experiments will provide the major data base on the effects of coal combustion conditions on pollutant control and particulate characteristics. Specific activities include:

1) A bench scale study to establish the role of combustion conditions on sulfur capture for a wide range of system designs and operating variables. Emphasis is placed on identifying the key variables for pollutant control optimization and providing guidance for specific system applications.

2) Small pilot scale experiments (1 to 10 x 10^6 Btu/hr) for low NO_x burner concepts for wall- and tangentially-fired boilers. Previous EPA work has identified NO_x control optimization, therefore these experiments will address SO_x capture and NO_x/SO_x control tradeoffs for selected conditions identified in the bench scale studies.

3) Ash characterization studies will be carried out on a bench scale apparatus at selected conditions. Emphasis will be placed on particulate characteristics including: slagging and fouling indices; particle size distribution and composition; properties related to control devices; and sorbent separation and residual activity for reuse.

4) To achieve high efficiency (i.e. 90%) SO_x capture for new systems will require an add-on device for incremental SO_x control. Small pilot apparatus (e.g. spray driers) will be used to evaluate use of residual sorbent activity with these devices.

5) Innovative concepts for high efficiency SO_x capture without add-on devices will also be evaluated. New combustion schemes afd/or heat transfer configurations which have the potential to accomplish this will be evaluated on a bench scale.

C. Prototype testing using full scale low NO_x burners will provide a basis for extrapolation of the development work to practical scale. The burners to be tested will include advanced burners for optimum NO_x and SO_x control for new systems and retrofit burners for moderate NO_x and optimum SO_x control for existing boilers. This testing will be carried out in the EPA's unique 150 x 10^6 Btu/hr large waterlube simulator facility and others as appropriate. The burners to be tested include:

1) Advanced experimental low NO_x burner prototypes, especially the EPA's distributed mixing burner concept, will be tested to optimize NO_x and SO_x control. When fitted to new systems this type burner should provide maximum possible pollutant reduction.

2) Commercial first and second generation low NO_x burners will be tested to determine optimum performance of off-the-shelf hardware. The manufacturer of each burner will participate in the testing and optimization of performance. This will provide direct experience for the manufacturer and a comparison to the performance of the optimum experimental system in C-1 above.

3) Internally staged burners, which can be easily fitted to existing boilers, will be tested in both experimental and commercial versions. It is expected that these burners will represent a compromise for NO_x control but may allow a significant cost saving for retrofit.

D. Process analysis will be performed to evaluate mechanical tradeoffs in system components and to analyze potential operability/reliability problems of the system. This activity is directed toward providing guidance for field application of the technology for both general and specific sites. Two types of activities are planned as follows:

1) A system study will evaluate mechanical aspects of the technology including: pulverization arrangements; boiler/burner design; radiant and convective heat transfer; particulate control devices; and sorbent recycle/SO_2 systems. A generic analysis will be performed for two types of boilers and cost estimates will be generated. This activity will pull together all aspects of the program into a general process approach.

2) In support of the process analysis, detailed time/temperature/-composition profiles will be established for representative boiler designs. This information is currently unavailable but may be critical to the final private sector decision on applicability to practical boilers.

E. Coordination will be emphasized throughout the program. Specific areas are as follows:

1) Industry will be involved by continuation of formal review through the technical panel meeting previously conducted under the low NO_x burner program.

2) EPRI and DOE will be briefed periodically on an informal basis.

3) The International Energy Agency Agreement on Combustion of Non-U.S. Coals will be used to coordinate with the member countries.

4) The informal exchange with German researchers will be continued.

VI. Recent Developments

Since the EPA work in 1978 several experimental programs have been initiated. The status of each of these is briefly described below:

A. The EPA has carried out a series of bench scale experiments directed at understanding the important mechanisms of sulfur capture. Results have shown that both thermal environmental and reaction stoichiometry are important factors. SO_x capture in excess of 70% at 2 times stoichiometric calcium have been achieved at selected conditions. The experiments are in an early stage and more work is needed.

In addition two large pilot scale low NO_x burners of the distributed mixing type have been tested. In both cases sulfur captures in excess of 50% have been achieved at a calcium stoichiometry of 2 in an experimental system simulating a large watertube boiler.

B. The most advanced effort to date is that of the German utility Rheinnisch-Westfailsches Electrizitatswerk (RWE). Limestone addition has been successfully tested on a 60 MW brown-coal fired utility boiler, achieving in excess of 50% capture. This work is not directly applicable to the U.S. situation because of the unusual characteristics of both the fuel and the boiler. Brown coal is even lower rank than U.S. lignite and has low heating value and sulfur content, and high moisture and alkali ash content. The boiler is a large relatively cold unit designed to accept a fuel with adverse fouling characteristics. While the experience provides valuable information, it bears no direct resemblence to the systems of concern in the U.S.

C. German boiler manufacturer, Steinmuller GmbH has run small pilot scale tests (6 x 10^6 Btu/hr) using a proprietary distributed mixing burner derived from data generated [illegible] earlier EPA study. At small scale SO_x removal in excess of 50% [illegible] achieved. Steinmuller is planning a limited trial on a low [illegible] utility boiler. The trial will use only 2 of the 24 burners [illegible] in due to limited simularity of the fuel and boiler design the [illegible] not be directly applicable to the U.S. situation.

VII. Summary

Based on the status of the testing to date the following comments can be made:

A. Even with limited testing, the target numbers for SO_x reduction have been achieved in several different experimental systems;

B. A systematic development and testing program is necessary to optimize the technology and to generalize for the full range of U.S. systems and fuels;

C. Particular attention must be given to operational and reliability aspects of practical boilers. Proper handling of the additional particulate loading is required.

D. While the EPA program does not allow field evaluation, generalization of the technology for application by industry is being emphasized, and will facilitate demonstration by the private sector.

Fluidized Bed Combustion (FBC) has the potential to reduce substantially SO2 emissions. This article from Environmental Science and Technology, reviews the development of FBC in the United Kingdom, where initial development centered on its use for utilities. Small and large industrial boilers can also incorporate FBC systems which the author describes in detail.

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The development of fluidized bed combustion

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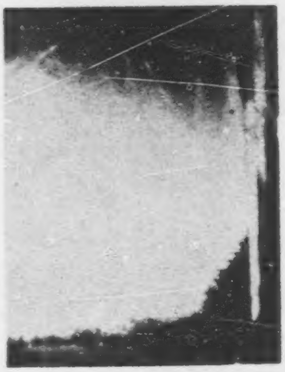
During the 1960s, most energy planners in the U.K. believed that the only major future use for coal would be in power generation, and that even this use would be limited by the planned increase in nuclear power and the availability of cheap oil. Despite the discouraging outlook, the U.K. National Coal Board (NCB) took the initiative during this period to develop a new coal-burning technology, fluidized bed combustion (FBC). The main objectives were to reduce the capital cost of coal-burning power stations and to facilitate the efficient use of lower grade coal with high or variable ash content.

It was soon found, however, that fluidized combustion would also allow a substantial reduction in the emission of sulfur dioxide, simply by feeding crushed limestone with the coal. The potentially low environmental impact of this new technology created considerable interest in the U.S., where, by 1970, it was becoming apparent that coal would have to be used to supplement diminishing oil and gas reserves. Interest in the U.K., however, remained low; oil was becoming available from the North Sea, and there was no legislation limiting sulfur dioxide emissions. From 1970 to 1973, most research and development of FBC in the U.K. was funded by U.S. agencies.

In 1974 the situation changed. Following a major study of the U.K. coal industry, the government authorized the NCB to implement a "Plan for Coal" that will increase annual output to 150 million tons. The additional coal is expected to displace oil and gas in industrial and commercial applications.



This new technology holds the potential for clean, efficient burning of coal in applications ranging from heating to electric generation. A British expert examines its development and current status

Coal crushing eliminated

In anticipation of this planned increase in coal use, the NCB gave fresh impetus to its development of FBC, concentrating on the industrial market. Although conventional coal-burning equipment is efficient and proven, it is expensive and offers a low level of amenity in comparison with oil and gas systems. Developing a modern coal-burning system which can be installed at a low capital cost is thus an essential step in increasing the direct use of coal; the alternative is the relatively inefficient production of synthetic liquid or gaseous fuel from coal.

Prior to 1973, the target of the FBC program had been coal-fired power station boilers. The fluidized combustion concept which evolved appeared attractive for these very large boilers, and subsequent progress in the U.K., U.S., and elsewhere is confirming its potential. The system is inappropriate for industrial boilers, however, because it requires crushing and drying the coal before firing, and use of a tall combustion chamber (about 4 m) and a deep fluidized bed—thereby necessitating a high-power air fan.

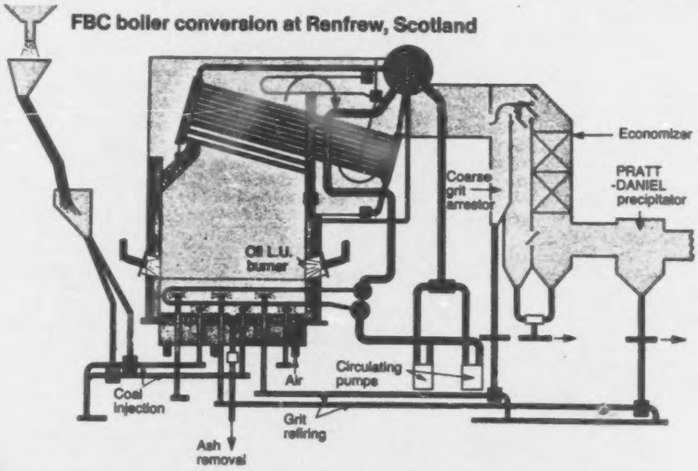
Analysis of the standard FBC system showed that its disadvantages were linked: The deep fluidized bed and tall combustion chamber were necessary to ensure efficient combustion of the finer coal particles inevitably produced by crushing. It was clear that elimination of coal crushing would lead to a more attractive system for industry and that the reasons for crushing had to be reconsidered.

The main reason for firing crushed coal is the need to maintain an effective fluidized bed in the combustion chamber. The burning coal, ash, and added limestone must remain suspended in the upward air flow, forming a violently churning layer of particles. During operation, the fluidized bed contains very little coal (<5%) and consists almost entirely of ash and limestone particles, which must be below about 5 mm in size for effective fluidization. Coal supplied to U.K. power stations contains stone from bands within the coal seams, having diameters up to the maximum coal size of 55 mm. At an early stage in the development of fluidized combustion, it was decided that crushing offered the simplest means of ensuring the correct size of ash (stone) particles in the fluidized bed. Crushing also ensured that the burning coal particles could be fluidized and thus well distributed through the bed.

But because the stone mined with coal has a higher density, it is relatively simple to separate by a process referred to as "washing." Coal washing in fact reduces transport costs and minimizes the quantity of ash for disposal; the resulting savings have made the use of washed coal economic for almost all industrial users. The washed coal, with an average sulfur content of only 1.35% and an ash content between 3% and 10%, is marketed as "singles" (25–13 mm) and "smalls" (13–0 mm). With the recognition that, unlike power station coal, these standard U.K. industrial coal grades contain few discrete stone particles, came the realization that the main reason for coal crushing no longer applied.

Combustion tests confirmed that the commercial coal grades could be burned by simply feeding, as supplied, to the surface of a fluidized bed of silica sand. As expected, the elimination of crushing brought about a substantial improvement in combustion; the target of 97% combustion efficiency was achieved when burning the singles grade coal in a fluidized bed 150-mm deep in a combustion chamber only 900 mm in height, without the usual complication of a secondary combustion stage. With the smalls grade coal, the performance in this compact high-intensity system was similar to that previously obtained with crushed coal in a deep bed and tall combustion chamber.

This variant of fluidized combustion is being used in a new generation of compact industrial boilers and furnaces. In the U.K., several demonstration/prototype boilers and hot gas furnaces are in daily use at industrial sites; within the next year, most larger boiler-manufacturing companies are expected to offer package boilers on a standard commercial basis. In the U.S., the Johnston Boiler Co., Ferrysburg, Mich., is already selling



FBC boiler conversion at Renfrew, Scotland

FBC in large-scale demonstration

The first large-scale demonstration of FBC was in the conversion of an existing water-tube boiler, using a design suitable for utilities and high-output industrial boilers. The conversion was carried out by Babcock Power Ltd. (BPL) at its Renfrew, Scotland, works. The technology was developed by the National Coal Board and British Petroleum, and was licensed by their jointly owned company, Combustion Systems Ltd.

The boiler is a cross-tube type, rated at 13.5 MW of saturated 25-bar steam when fired with a spreader stoker. The FBC system consists of a bed 3.1 m by 3.1 m, with horizontal immersed hairpin tubes designed for evaporation with forced water circulation.

The boiler was commissioned in 1975 and has operated satisfactorily for more than 5000 hours. Studies of light-up; load control; uniformity of bed temperature; effect of fuel size, number of feed points, and fluidizing velocity on unburned carbon carry-over; heat transfer to immersed tubes; corrosion and erosion of tubes; sulfur retention by limestone; and NO_x emissions have been carried out on this unit. A wide range of coals have been burned with efficiencies at least as good as those obtained with the original stoker. Retention of 90% of the sulfur was attained when burning a 3.5%-sulfur coal. NO_x emissions were found to be well within present limits for new coal-fired boilers in the U.S.

The first commercial application of the FBC system that was demonstrated at Renfrew is at the Central Ohio Psychiatric Hospital in Columbus. Babcock Contractors Inc. (BCI) of Pittsburgh (a subsidiary of Babcock Power Ltd.), in collaboration with the Riley Stoker Corp., converted an existing 20-MW boiler. The unit, which will be commissioned shortly, will burn high-sulfur Ohio coal.

Currently, Fluidized Combustion Contractors Ltd., a company recently formed by Babcock Power Ltd. and British Petroleum, is the principal U.K. organization for engineering FBC systems for utility and large industrial applications. They will be closely involved in carrying out a contract recently awarded to BCI by the Tennessee Valley Authority for the detailed design of a 200-MW (electric output) boiler to burn West Kentucky coal containing 4.5% sulfur with 90% sulfur retention.

boilers incorporating this technology under license. The recently commissioned 30-MW boiler at Georgetown University, designed and manufactured by Foster-Wheeler, also makes use of the advantages offered by eliminating coal crushing.

Boiler design for FBC

Fluidized combustion is more than just a new technique for burning coal with minimum environmental impact. It also provides boiler designers with an opportunity to make significant cost savings.

Steam-raising tubes can be immersed within the bed of turbulent red-hot particles to obtain high heat-transfer rates. Extensive trials have demonstrated that, contrary to early fears, corrosion and erosion of these tubes are minimal. However, a boiler with tubes immersed in the fluidized bed presents novel problems of start-up and control, and these must be solved by a combination of boiler design and control strategy. Various approaches are possible and unfortunately there is no obviously "best" design; consequently, there is considerable variety in the designs of first-generation FBC boilers.

The air supplied to the fluidized bed is heated to the operating temperature of about 900 °C. The heat required for this is approximately 40% of the heat release which would be obtained if coal were burned at the theoretical maximum rate for the air supply. The burning rate is limited to this 40% level unless tubes are immersed in the bed to remove additional heat. The *optimum* rate, however, is determined by the need to ensure efficient combustion and prevent unacceptable emissions of unburnt volatiles and carbon monoxide. This is normally about 80% of the theoretical maximum for the air supply; in other words, efficient combustion requires 25% excess air. At this optimum rate, about half of the heat release is transferred to the tubes and the remainder is removed by the hot combustion gases to be recovered in the convection section of the boiler. Having too few tubes in the bed would necessitate a lower burning rate, which would reduce total power output; having to heat a greater proportion of excess, unused air would at the same time reduce efficiency. On the other hand, having too many tubes would necessitate a burning rate greater than optimal, causing incomplete combustion.

These considerations apply throughout the output range of a boiler, and it is necessary to incorporate design features which ensure that the direct heat transfer to tubes in the bed is maintained close to, but does not exceed, 50% of the heat release. This is usually achieved by varying the bed temperature within its operational range and by making provision to reduce the area of tubing in the active bed as output is lowered.

Start-up of the FBC boiler involves heating the inert bed to about 500 °C, using a premium fuel, after which coal feeding can be initiated. The amount of premium fuel required can be minimized by limiting heat transfer to the tubes in the bed during start-up. Heating techniques include firing oil burners above the bed, passing premixed air and gas through the bed to give a flame at the surface, and fluidizing with hot gas from an external burner.

Oil-to-coal conversion

Developments in oil firing over the past 20 years have led to substantially more compact boilers than those for conventional coal stokers. Also, in the U.K., horizontal fire-tube construction has almost entirely replaced the more expensive water-tube construction in boilers with outputs up to 20 MW (60 000 lb/h of steam).

In the early stages of the NCB program, it was considered that there would be considerable market potential for an FBC firing system for fire tube boilers, particularly if this would permit the conversion of existing boilers from oil to coal with minimum reduction in output. In 1975, the NCB started a collaborative development towards this objective with Parkinson Cowan GWB Ltd., using one of their Vekos Powermaster boilers, rated at 1.2 MW. The fire tube of a conventional boiler is less than ideal for fluidized combustion since it is intended for horizontal gas flow rather than vertical, but, following a number of modifications, a unit was operated satisfactorily for extended periods. The output achieved was similar to that of a conventional stoker; work is continuing on alternative designs for higher output. However, recent design studies for fire-tube boiler conversions from oil indicate that the market potential is less than expected because of the difficulty of installing coal storage and handling systems, grit arrestors, and ash systems into existing boiler houses. Consequently, the main market potential for FBC systems in fire tube boilers is now seen to be in new boilers, initially for the same output range as stoker-fired units, and in existing boilers originally installed for coal but since converted to oil.

In an independent program, Northern Engineering Industries Ltd. has developed an FBC system with an output similar to that of a stoker, using a 1-MW boiler. This company, in which the U.S. firm Combustion Engineering is a shareholder, is the U.K. market leader in fire tube boilers through its subsidiary N.E.I. Thompson Cochran Ltd. A 5-MW fire tube boiler has recently been commissioned and will be used to prove the FBC system prior to commercial manufacture of a range of boilers up to 6-MW output.

After initial investigations supported by the NCB, Energy Equipment Ltd. has developed a system which enhances combustion above the fluidized bed by recirculating flue gas through the bed and secondary air injection. Although intended primarily for fire tube boilers, the system has been used to convert a 16.7-MW water tube boiler from stoker firing. Energy Equipment has also been given a contract by the NCB to provide FBC boilers for a new mine.

New FBC boilers

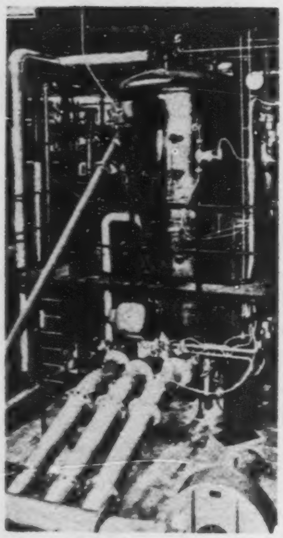
The development of new purpose-designed boilers for FBC offers the major advantage that the combustion chamber can be arranged for vertical gas flow and the correct amount of tubing can be installed in the bed in a configuration suited to achieving simple start-up and effective load control. Although this is readily achieved with water tube boilers, they are expensive to manufacture, and so there is a strong incentive to use shell-type boilers in order to compete more effectively with low-cost oil-fired package units. For smaller outputs, up to about 5 MW, vertical shell boilers are potentially attractive, but for higher outputs it is necessary to devise some method of incorporating the vertical combustion chamber into a horizontal boiler shell.

Two vertical shell boilers, one for steam and the other for hot water, have been designed by the NCB and operated at commercial sites. Both were installed in 1977, were operated through the 1977–78 and 1978–79 heating seasons with NCB supervision, and are expected to continue in commercial operation. The steam boiler, at a factory in Lancashire (see photo), incorporates automatic start-up and load-following systems. It successfully met the heat demand of up to 2 MW throughout the 1978–79 heating season, during which it was the only operational boiler at the site. The hot water boiler, at a market garden in Hereford, has a rated output of 3 MW, but has only occasionally been required to operate above 1 MW. It has been modified to facilitate control at low outputs.

With the experience gained from these two prototypes, the NCB is collaborating with Vosper Thornycroft (U.K.) Ltd. to design a range of vertical shell boilers. The first of these boilers, rated at 4.5 MW, is being installed at a district heating scheme near London and will be commissioned early in 1980. Five similar boilers are to be installed at various sites during 1980.

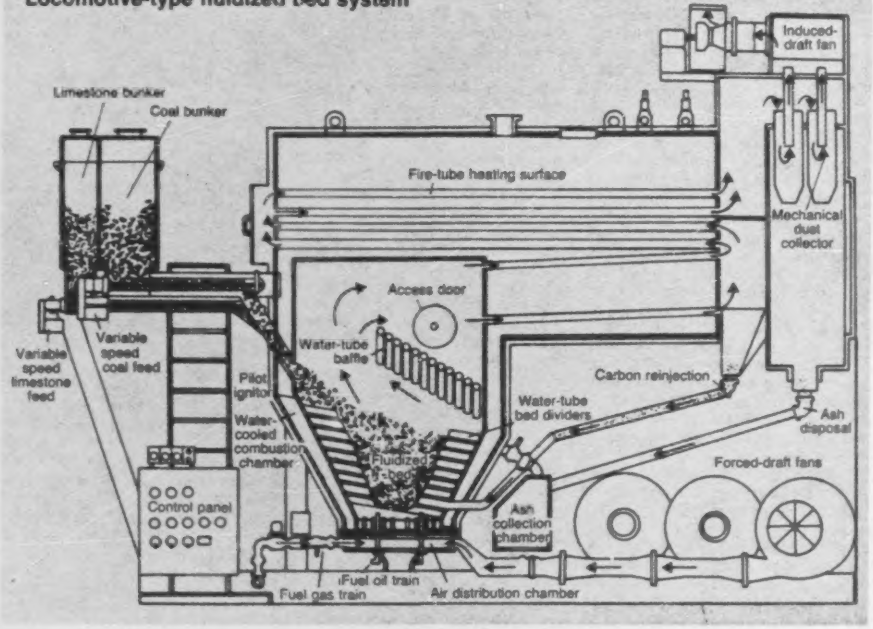
The traditional method of incorporating a vertical combustion chamber into a horizontal shell boiler is to use a locomotive boiler configuration. To demonstrate this concept for FBC packaged boilers, the NCB has designed and installed, at a factory in the Midlands, a 10-MW unit. This is basically a development boiler, with several novel features in the FBC system, but it is now reaching the stage where it will reliably meet the factory load. So far there is no commercial manufacturer of this type of boiler in the U.K., but in the U.S., the Johnston Boiler Company (JBC) has adopted a locomotive-type design for its FBC boiler range (Figure 1). Using U.K. technology licensed through CSL, JBC designed and manufactured a 3-MW (10 000 lb/h) prototype, which was put into operation in September 1977. This has successfully fired natural gas, distillate and residual fuel oil, and numerous bituminous coals, as well as waste materials including logged wood, sawdust, shredded rubber, waste oil, and paper sludge. JBC is now offering its "Fluidfire" multifuel boilers for outputs up to 15 MW on a standard commercial basis. The first production unit, a boiler of 40 000-lb/h output for Central Soya, has been shop tested and should be in operation by February 1980. A second unit of 20 000-lb/h output is scheduled for installation at another site a month later.

The alternative approach of developing packaged water-tube boilers for FBC has not been ignored in the U.K. and is being followed by Stone-Platt Fluidfire. This company, which pioneered the use of gas-fired fluidized beds for metallurgical heat treatment, has developed a fluidized combustion method for coal and other solid fuels which incorporates circulation of the bed particles to improve combustion and sulfur retention. They have provided small (300 kW) boilers for in-

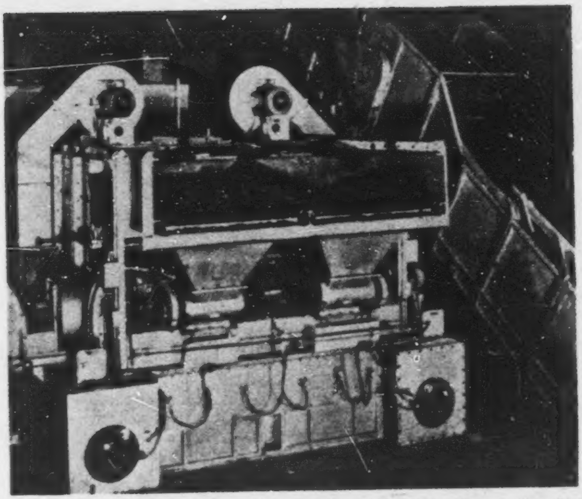


Vertical shell boiler. *This prototype 2-MW steam unit has operated since 1977 at a factory in Lancashire, U.K.*

Locomotive-type fluidized bed system



Induced-draft fan
Limestone bunker
Coal bunker
Fire-tube heating surface
Mechanical dust collector
Access door
Variable speed limestone feed
Variable speed coal feed
Water-tube baffle
Pilot ignitor
Water-cooled combustion chamber
Fluidized bed
Carbon reinjection
Water-tube bed dividers
Ash disposal
Forced-draft fans
Control panel
Ash collection chamber
Fuel oil train
Fuel gas train
Air distribution chamber



Grass dryer. *FBC furnaces can be used to supply hot gas for drying plants, such as this grass dryer producing cattle feed*

vestigation by Virginia Polytechnic Institute and General Motors, and will shortly be supplying a 3.75-MW unit to a factory in northern England with U.K. government support. Stone-Platt has recently acquired a majority interest in the Johnston Boiler Company, strengthening the FBC expertise and marketing potential of both organizations.

New 30-MW boiler

The National Coal Board, in cooperation with ME Boilers Ltd., has designed and built a 30-MW FBC coil-type boiler to produce superheated steam at 43 bar. This is at the steel works in Sheffield and will burn commercially available coal sized 18–0 mm, with start-up by premixed air and coke oven gas. It is to achieve a turndown ratio of 6:1 by having a circular bed divided into six independent compartments. The requirement to provide a rapid increase in steam supply from low output will be met by maintaining at least three of the compartments on stand-by at a temperature from which coal combustion can be ignited, the beds expanding into the tube bank on fluidization. A single compartment has already been fired on gas and it is planned to bring the boiler into operation in early 1980.

Other applications

Drying processes are major users of heat in industry and agriculture, and the firing of dryers was identified as a potential application for FBC in 1973. Development work has so far concentrated on applications where the hot combustion gases are in direct contact with the product. This restricts coal firing to products in which a small amount of ash contamination is acceptable, but there are many of these. In the FBC furnace, all of the heat released by combustion is removed from the bed by the combustion gases (plus substantial air), giving hot gas at up to 950 °C. The commercial application of FBC to the firing of grass dryers for producing cattle feed was first demonstrated in 1974; there are now five 5-MW FBC grass dryers in commercial operation, and further units are to be installed in 1980. The manufacturer, G. P. Worsley Ltd., is collaborating with the NCB to extend the output range and apply the furnaces to other processes. A 15-MW furnace has recently been commissioned to dry clay in cement manufacture, and a unit has been designed for drying and heating roadstone. Work is now being initiated to investigate methods of adopting FBC technology to produce dust-free and uncontaminated hot gas so that almost any product may be dried with hot gas from coal.

Another potential use of FBC is in the incineration of waste products. Such applications range from combustion of high calorific value materials such as wood waste, for which almost standard FBC boilers can be used, to the incineration of waste with a low heat content. In the U.K., the NCB has demonstrated the use of FBC to dispose of coal washery tailings, which is a slime comprising 50% water, 35% ash, and 15% coal. This technology has been licensed through CSL to Heenan Environmental Systems, which designed and built an incinerator for sewage sludge at Caernarvon, North Wales, that has been in operation for about two years.

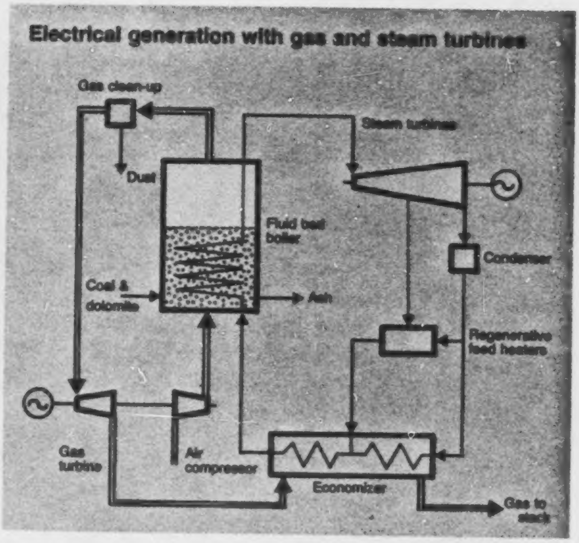
Improving generation efficiency

Energy research has for many years been concerned with improving the efficiency of electricity production. At present only about 38% of the fossil fuel energy consumed by a power plant is converted to electricity; the remainder is discharged to the atmosphere through cooling towers. An accepted route to higher power-generating efficiency is to incorporate a gas turbine with a steam turbine in a combined thermodynamic cycle; but this route has been precluded until now by the high cost of premium fuels required by gas turbines. It was recognized in the late 1960s, however, that FBC could be used for a coal-fired gas turbine that would achieve high efficiency. The relatively low combustion temperature of FBC means that the alkali compounds harmful to turbine blades are not volatilized; also, the ash particles in the gases are not vitrified, and are consequently less erosive than those from other coal combustion systems.

The first stage in the development of this advanced concept was to demonstrate that a coal-burning FBC unit could operate at elevated pressure. Work started in 1968 at the British Coal Utilization Research Association (now the NCB's Coal Utilization Research Laboratories), and in 1969 a 2-MW pilot plant was commissioned. In a major research program over 10 years (with funding from the U.S. Department of Energy since 1972), this plant has shown not only that the gas quality is likely to be acceptable for turbines, but also that operation at elevated pressure improves combustion efficiency and sulfur retention while lowering emission of NO_x. The next major step will be the commissioning in 1980 of the International Energy Agency 80-MW (thermal) test facility at Grimethorpe in the U.K. This facility, with a bed cross section of 2 m by 2 m and the capability of operating at pressures up to 12 bar, is jointly financed by the U.S., the U.K., and West Germany.

Although tests with static turbine blade cascades continue to give fa-

Gas turbine system. *The lower combustion temperatures possible with FBC result in reduced emissions of compounds harmful to turbine blades. This 2-MW pilot combustor, shown above its pressure vessel, has been used in a 10-year program aimed at demonstrating the feasibility of combining an FBC boiler with a gas turbine to improve electric generation efficiency*



Electrical generation with gas and steam turbines

vorable predictions of acceptable blade life, there will be no certainty of success until an industrial-sized gas turbine has run for many thousands of hours. However, the concept promises a fuel saving of at least 6% for utility power generation and is also attractive for the large-scale industrial market where heat and power are required.

Environmental impact

The U.K. is fortunate in that its coal reserves have a low sulfur content, with present production averaging only about 1.5% sulfur. Use of coal in place of oil will thus have little impact on the presently acceptable atmospheric SO_2 concentrations in urban and rural areas, which have been achieved through a policy of dispersal which relates stack height to fuel sulfur content and thermal output. It is expected that air quality standards which are at present being drafted by the European Economic Community (EEC) will not require a change in U.K. policy, although sulfur retention legislation will be required and is being introduced in countries where the coal has a higher sulfur content, notably Germany. It appears, however, that SO_2 concentrations which are acceptable in the EEC countries might be causing a gradual increase in acidity of the already acid soil and inland waters of Scandanavian countries. If this is substantiated, there could be pressures to restrict SO_2 emission throughout Europe, but only in this case is it likely that the potential of FBC to retain sulfur will be used in the U.K.

Matching the application

Over the past decade, FBC has evolved from a design concept for utility boilers to a system which can be considered for a wide variety of applications. Already, a range of boilers, firing units for dryers, and incinerators are commercially available in the U.K. The development of FBC at elevated pressure is nearing the point at which a large gas turbine can demonstrate the high efficiency achievable in combined cycle power generation. Coupled with this variety of applications is the wide range of fuels which can be used (including coal of every type and quality as well as low-grade solid, liquid, and gaseous wastes), and the potential for meeting the wide variety of national environmental regulations. The range of FBC designs and operating concepts is equally wide, and it may not be possible to transfer a design developed for a specific application to other situations in another country. However, to select the best design approach for any application, it is essential to take account of the expertise available in every country. In particular, the progress made in the U.K., operating coal-burning boilers and furnaces in a commercial environment, complements the work in the U.S., where the use of high-sulfur coal necessitates greater attention to emission control.

Acknowledgments

The author wishes to thank the National Coal Board for permission to publish this article, but the views expressed are his own and do not necessarily reflect those of the NCB. Acknowledgment is made of the comments and suggestions made on the first draft by the author's colleague, Mr. D. C. Davidson.

Additional reading

H. R. Hoy and W. G. Kaye, "Work by the NCB on the development of atmospheric and pressurized fluidized bed combustion," *J. Inst. of Energy* **52**, no. 411 (June 1979) and other papers in this issue.

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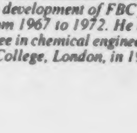

John Highley *is responsible for industrial coal burning and ancillary equipment projects at the National Coal Board's Coal Research Establishment. He was a member of the NCB research team that pioneered the development of FBC for utility boilers from 1967 to 1972. He received a B.Sc. degree in chemical engineering from Imperial College, London, in 1964.*

This Tennessee Valley Authority (TVA) report reviews atmospheric fluidized bed combustion (AFBC), its operation, advantages, costs, and commercial availability timetable. Cost and performance data are presented on page 643.

TECHNOLOGY DEVELOPMENT
and DEMONSTRATION

Tennessee
Valley
Authority

November 1, 1982

TVA
Tennessee
Valley
Authority

TVA is a corporate agency of the United States Government. It was established by Act of Congress in 1933 to develop the Tennessee River system and to assist in the development of other resources of the Tennessee Valley and adjoining areas.

The production and sale of electric power are part of TVA's resource development program. TVA supplies power at wholesale to 160 municipal and cooperative distributors and one privately owned electric system which in turn distribute power to over 2.8 million customers in parts of seven States. TVA also supplies power directly to 50 industrial customers with large or unusual power requirements and several federal, nuclear, aerospace, and military installations.

AFBC

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Introduction

This brochure describes the Tennessee Valley Authority's (TVA) experience with atmospheric fluidized bed combustion (AFBC) and discusses the need for a demonstration plant to show the practicality of AFBC before potential users undertake a commercial-scale project. AFBC is an important alternative to conventional pulverized coal (PC) burning boilers with flue gas desulfurization (FGD) for electric power generation because of its projected lower capital and operating costs as well as its environmental acceptability.

TVA has committed more than $87 million in research and development in AFBC during the past six years and is willing to commit additional funds in the future to carry the development forward. This total expenditure includes related research and development, design and construction of a 20-MW pilot plant in Paducah, Kentucky, which was dedicated in August 1982 and has been operating since then. The pilot plant is providing valuable data for the design of the demonstration plant. In turn, the demonstration plant is needed to prove the technology and economic feasibility of commercial utilization of AFBC by utilities. This includes providing answers to key questions such as capital and operating costs, reliability and emissions levels.

TVA proposes to build a 160-MW AFBC add-on boiler facility at its Shawnee Steam Plant in Paducah, Kentucky. Conceptual studies for the boiler configuration, related auxiliaries and the balance of the plant (BOP), as well as preliminary cost estimates, have been prepared. Engineering and cost studies currently are being refined. Detailed design of the demonstration plant will be initiated in the summer of 1984. The plant is scheduled to commence operation during 1989. Testing will be carried out during the first few years of operation so that design data will be available during the early 1990s to utilities considering building commercial AFBC stand-alone plants and replacement facilities which would come on-line during the mid-to-late-1990s.

The preliminary cost estimate for a 160-MW AFBC add-on demonstration plant at Shawnee is approximately $200 million. TVA will commit funds to this project, but because of the substantial cost of an AFBC demonstration plant and the benefits that can accrue to many utilities, coal companies, equipment vendors and government agencies, TVA proposes that a funding consortium be established whose members would share the benefits as well as the costs.

Further details about the demonstration plant project can be provided by the TVA staff. Please address requests to:

Project Manager
Atmospheric Fluidized Bed Combustion
Demonstration Project
Division of Energy Demonstrations and
Technology
Tennessee Valley Authority
1020 Chestnut Street, Tower 2
Chattanooga, Tennessee 37401

I. Description of Atmospheric Fluidized Bed Combustion (AFBC) Process

A. OVERVIEW OF THE PROCESS

Atmospheric Fluidized Bed Combustion (AFBC) is a process of burning coal in a bed of limestone particles that are held in suspension by the upward flow of combustion air entering a chamber through a perforated air distribution grid below a fluidized bed. The major elements of a fluidized bed boiler are shown in Figure 1 and briefly described below.

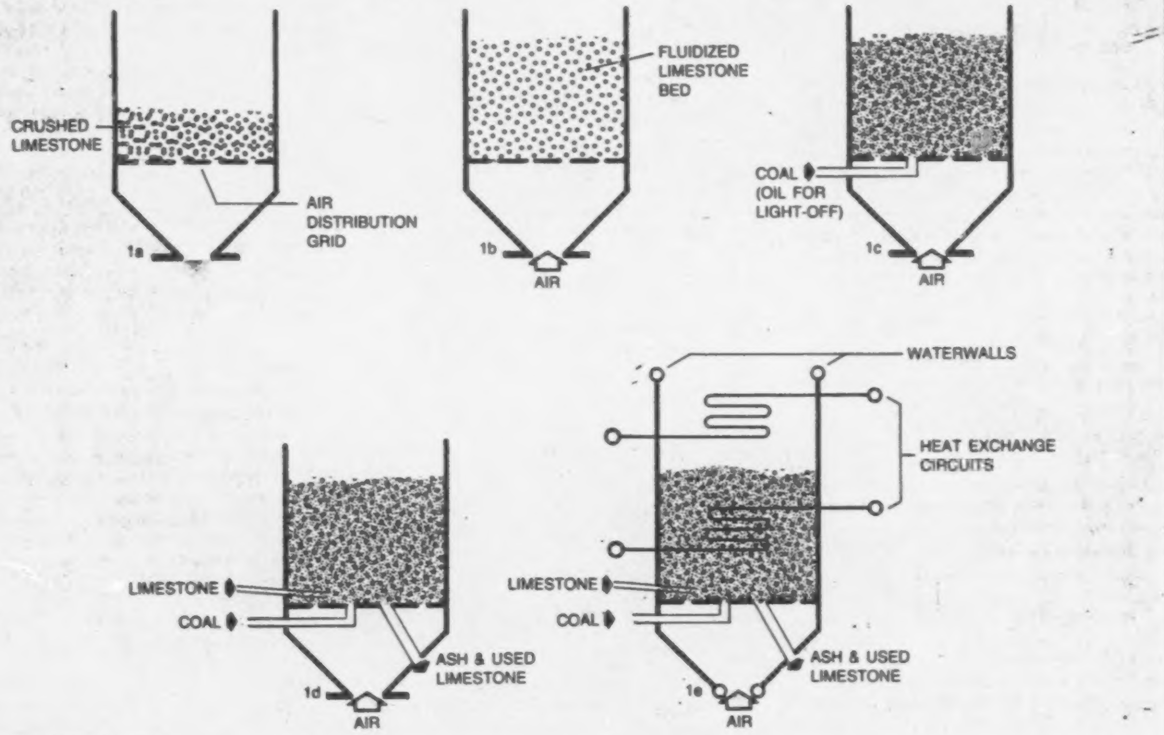
First, crushed limestone is placed in the bed on the air distribution grid (Figure 1a). The bed is then fluidized by injecting air through the grid to form an expanded, suspended mass that closely resembles a boiling liquid (Figure 1b). A fire fed by oil or other fuel is injected into the bed to heat the material (usually coal) to ignition temperature (Figure 1c). During the combustion process, limestone is added with the coal which greatly reduces sulfur dioxide emissions (Figure 1d).

During operation, coal and limestone are fed continuously into the bed. The limestone reacts with the sulfur dioxide released from the burning coal and prevents the sulfur dioxide from escaping into the atmosphere. Steam is produced in tubing located in the fluidized bed and above the bed (Figure 1e). This steam then flows to a turbine generator to produce electricity. Coal ash and spent limestone are drained from the bottom of the bed. This dry residue can be disposed of by conventional methods; however, it is potentially usable as an agricultural supplement, road base filler, or cement additive.

B. BACKGROUND

The origin of fluidized bed combustion is generally traced to the Winkler gas generator developed during the early 1920s in Germany. Fluidized beds have been used in many chemical processes involving a reaction between a solid and a gas or between two gases where a solid catalyst is used to promote the reaction. It has been widely used in the steel industry for ore roasting. In the 1950s, fluidized beds were being commercially used for the incineration of municipal sewage sludge. In recent years there has been significant research and development work in the combustion of coal in a fluidized bed for both industrial and utility operations.

FIGURE 1
ELEMENTS OF A FLUID BED BOILER


CRUSHED LIMESTONE
AIR DISTRIBUTION GRID
1a
FLUIDIZED LIMESTONE BED
1b
AIR
COAL
(OIL FOR LIGHT-OFF)
1c
AIR
WATERWALLS
HEAT EXCHANGE CIRCUITS
LIMESTONE
COAL
ASH & USED LIMESTONE
1d
AIR
LIMESTONE
COAL
ASH & USED LIMESTONE
1e
AIR

Source: Tennessee Valley Authority

C. COMPONENTS OF THE FLUIDIZED BED SYSTEM

1. Introduction

The first AFBC designs which were developed more than sixty years ago consisted of an existing boiler coupled to a combustion chamber modified to accommodate the fluidized bed. These early units were the simplest type of fluidized bed -- the so-called bubbling bed. From the experience gained in their operation, designers added more sophisticated features to the basic process to improve performance. In recent years, the combination of improved performance, lower capital and operating costs and special advantages, such as reduction of sulfur dioxide and nitric oxide emissions, have increased interest in AFBC, especially for utility applications.

Two questions that generally are asked about AFBC include:

- How does AFBC differ from conventional combustion and how does this account for greater emissions control and fuel flexibility?
- What is the configuration of an AFBC chamber and how is the heat transfer surface incorporated in the bubbling bed?

The first question is discussed in Section 2 below and the second is discussed in Section 3.

2. Fluidization[1]

In a fluidized bed, solid, liquid, and/or gaseous fuel together with inert material — for example, sand, silica, alumina, or ash from the fuel — are kept suspended in the lower section of a combustion chamber through the action of fluidizing air distributed below the bed. Fluidization promotes the turbulent mixing conditions required for good combustion, the extent of which depends on the size of the particles, the bed temperature, and the fluidizing velocity.

The efficiency of combustion depends on three primary variables: time, temperature, and turbulence. The turbulence promoted by fluidizing the bed makes the entire mass of solids behave like a liquid. The resulting improvement in mixing allows generation of heat at a substantially lower and more uniformly distributed temperature than in conventional boilers — typically 1500-1600°F. A properly sized combustion chamber will release heat at an equivalent level to a conventional boiler, but at lower temperature without any theoretical loss in efficiency.

This fluidizing mechanism offers several advantages over conventional combustion:

- less volatilization of alkali compounds — a phenomenon known to cause deposits on boiler-tube surfaces;
- lower potential for local hot spots on boiler or shell surfaces;
- relative insensitivity to the quantity and nature of the ash in the fuel; the low combustion temperatures prevent vitrification of the ash particles, causing them to be less abrasive than ash from pulverized coal (PC)-fired units; and
- smaller furnace volume. Extra furnace volume is not needed to allow the ash to cool below its softening temperature before flowing through convective heat-transfer surfaces.

The last two factors are the essential ingredients in the fuel-flexibility edge attributed to AFBC boilers.

The combustion process also provides emissions control. The operating temperature is well below the thermal-NO_x formation point. By including a suitable sorbent as part of the bed material — such as limestone or dolomite — the SO_2 released during combustion can be absorbed, eliminating the need for downstream FGD.

Two reactions take place to accomplish SO_2 removal: limestone is calcined to produce calcium oxide and this then combines with SO_2 to form calcium sulfate. Studies show that the calcination reaction absorbs heat, while the SO_2 reaction releases heat. The two tend to be thermally balanced in the 1500-1600°F temperature range. When they are not, SO_2 removal efficiency deteriorates. Thus, there are performance tradeoffs in maintaining high SO_2 removal and high combustion efficiency.

3. Heat Transfer

To control temperature in the bubbling-bed design, a suitable heat sink is provided in the combustion area. This generally is in-bed boiler tubes, and/or waterwall enclosures. The tubes are usually used to generate and superheat steam. An alternative is to use air as the heat-transfer medium. It can then be used to drive a gas turbine or fulfill process-air requirements.

Steam or air output is controlled by manipulating the principal bed parameters: height, temperature, inventory, and superficial gas velocity. This requires reducing the heat flux in a large, well-distributed high-temperature area. This is usually accomplished in bubbling beds by making use of a compartmentalized arrangement. By reducing the fluidization velocity in one section of the bed, less heat transfer to the in-bed tubes and the waterwalls will occur. When the air is shut off completely, the heat transfer in that area is significantly reduced. This approach is called slumping the bed. Reducing the amount of bed material accomplishes the same thing. Also, fuel feed can be regulated as in a conventional boiler. Each of these methods provides a limited amount of turndown and, in practice, a combination of methods could be required.

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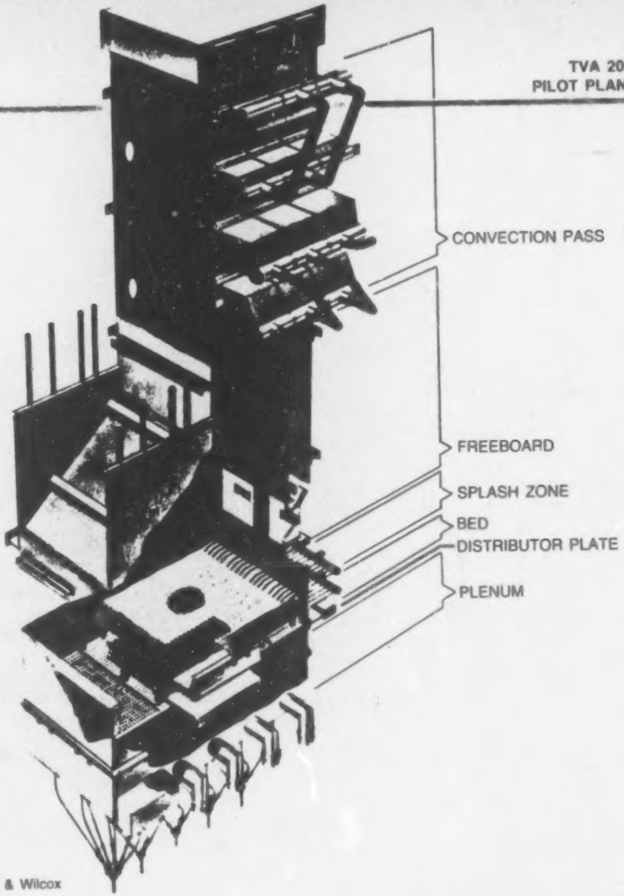
[1] This discussion is adapted from "Fluidized Bed Boilers" by Jason Makansi and Bob Schwiger in *Power*, August 1982.

To ensure good combustion characteristics, reacted limestone and bottom ash must be removed periodically. The bed material is hot, and equipment for this requires careful construction in order to recover thermal energy. In many cases, feedwater heating will recover heat and cool the ash-collection system.

Above the bed, the hot combustion gases are cooled in the convection passes in the freeboard area. The freeboard section also improves combustion efficiency by allowing elutriated particles to fall back to the bed and ensure more complete carbon burnout.

An example of an AFBC plant is shown in Figure 2. This identifies the principal components of TVA's 20-MW pilot plant which is currently operating in Paducah, Kentucky.

FIGURE 2
TVA 20MW AFBC
PILOT PLANT BOILER



Source: Babcock & Wilcox

II. Advantages of Atmospheric Fluidized Bed Combustion

A. INTRODUCTION

Atmospheric Fluidized Bed Combustion (AFBC) is of special interest to owners and operators of fossil-fueled electric generating plants for two basic reasons:

- air quality emission standards can be met when burning lower-cost, high-sulfur, high-ash, coals without expensive FGD systems, and
- the capital and operating costs of AFBC processes are projected to be lower than for PC systems with FGD.

It is estimated that the AFBC process can generate electricity at a rate that is 10 to 15 percent lower[1] than for conventional PC with FGD with scrubbers (see Table 1). The economic advantages of the AFBC process will help to make it possible for the utility industry, as well as many other industries, to return to using high-sulfur coal.

In the future, AFBC technology may be used by the utility industry for new capacity, replacement capacity, and substitution capacity for oil and natural gas units. The national significance of AFBC technology is indicated by the fact that in the United States there are approximately 1,295 coal-fired units which were commissioned before 1960, ranging in size from 4 to 500 MW. These units have a total capacity of 98,500 MW. As these units age, replacement or refurbishment will be required. Commercially available AFBC technology could be used to replace such units in the mid-to-late-1990s. In addition, approximately 223,000 MW of new coal-fired capacity is estimated to be needed during the 1990 to 2000 time period.

B. GENERAL ADVANTAGES OF THE COMBUSTION PROCESS

The combustion process in an AFBC occurs in the presence of limestone, which reacts with the sulfur dioxide (SO_2) formed during combustion to produce calcium sulfate — a dry, solid material, which can be easily disposed of or potentially used for agricultural purposes. Thus, the process offers the potential for combustion of coal in compliance with the Federal Environmental Protection Agency's limitation on SO_2 emissions without the need for a FGD system. The combustion temperature is maintained below the softening temperature of the ash so the fuel ash remains in a solid form throughout the process. The design of the AFBC boiler is, therefore, much less dependent on the fuel ash characteristics than is the design of a conventional PC boiler with FGD. Also, the low combustion temperature significantly reduces the formation of nitric oxides (NO_x). The size and cost of the AFBC boiler is less dependent on fuel characteristics than PC boilers with FGD and therefore offers the potential for more cost-effective utilization of low-grade fuels than is possible with conventional PC boilers with FGD.

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[1] This is based solely on lower capital costs and does not take into account the lower cost of high sulfur coal.

TABLE 1
COST AND PERFORMANCE DATA FOR COAL-BASED GENERATION TECHNOLOGIES

	CONVENTIONAL PULVERIZED COAL			AFBC	GCC	
	SUBCRITICAL 2400 PSI FGD	SUPER-CRITICAL 3500 PSI FGD	LOW HEAT RATE 4500 PSI/ CHIYODA FGD	AFBC	TEXACO 2000°F	TEXACO 2600°F
FIRST COMMERCIAL SERVICE	CURRENT	CURRENT	1992	1993	1991	2000
NET HEAT RATE (BTU/KWH)	9860	9480	8420	9640	9400	8465
NET EFFICIENCY (%)	34.6	36.1	40.5	35.3	36.1	40.3
$/KW (IN SERVICE JANUARY 1981)	995	1000	1105	820	1030	965
LEVELIZED MILLS/KWH (1981-2010)	85	81	78	77	81	79

NOTES: AFBC MEANS ATMOSPHERIC FLUIDIZED BED COMBUSTION.
GCC MEANS GASIFICATION COMBINED CYCLE.
FGD MEANS FLUE GAS DESULFURIZATION.

Source: Tennessee Valley Authority, Division of Energy Demonstrations and Technology (August 1982) and the Electric Power Research Institute 1981 Technical Assessment Guide.

C. SPECIFIC ADVANTAGES

Some of the specific advantages of AFBC compared to conventional PC boilers with FGD are listed below:

Lower Cost

- AFBC has lower capital costs than PC boiler with FGD because of simpler design.
- AFBC has lower operating costs because FGD is not needed when burning lower-cost, high-sulfur, high-ash coals.

Environmental Advantages

- High-sulfur (eastern) coal can be burned and SO_2 air quality standards can be met without the use of FGD.
- NO_x emissions are lower than from conventional systems primarily because of lower temperature operation.
- Dry solid wastes (gypsum/ash) are easier to dispose of than the wet scrubber sludge generated in conventional systems.

Greater Efficiency and Flexibility

- Heat transfer is more efficient because the fluid-bed particles allow the heat transfer to occur primarily by conduction rather than by convection.
- Slagging problems are nonexistent because of the lower temperature operation of AFBC.
- There is more flexibility in the types of fuels burned, such as low- and high-sulfur coals and lower grades of coal.

Results of the first few months operations of TVA's 20-MW pilot plant in Paducah are encouraging.

III. Tennessee Valley Authority's Involvement in Atmospheric Fluidized Bed Combustion Technology and Development

A. INTRODUCTION

Since the mid-1970s, TVA has been pursuing the development of AFBC technology for utility applications. In 1976, TVA's Board of Directors approved Phase I of a program leading to the design and construction of a utility AFBC power plant. Contracts were awarded to three major boiler manufacturers to obtain conceptual designs and cost estimates for a 200-MW AFBC demonstration boiler and an 800-MW commercial size AFBC boiler as well as to identify research and development issues associated with this technology.

Using these studies as a basis, an economic analysis was performed. The results indicated that, in comparison to a conventional PC plant with FGD, an AFBC power plant can be constructed at a capital cost which is 10 to 15 percent lower than a conventional plant. Siting studies were conducted for the selection of a suitable location in the Tennessee Valley region for a 200-MW demonstration plant. In the early stages of the project, it was recognized that a pilot plant, which could be operated under utility conditions, was needed in order to confidently design a demonstration plant.

B. PILOT PLANT

In 1979, TVA awarded a contract to Babcock & Wilcox with a subcontract to Stone & Webster for the design and construction of a 20-MW AFBC pilot plant at Paducah, Kentucky. This $68 million plant was completed on schedule during the first quarter of 1982 (see Figure 3). It has undergone extensive shakedown testing, and a joint TVA/Electric Power Research Institute (EPRI) four-year test program was initiated during the summer of 1982. Goals of the pilot plant are to:

- prove that efficient and cost-effective performance of an AFBC boiler and auxiliary systems can be obtained under utility operating conditions and to facilitate the rapid development of this technology;
- develop and prove-out operating control schemes required for reliable utility performance relative to variations in power demand (such as load-following operations);
- test, compare, develop, evaluate, and select large-scale auxiliary components and AFBC facility design configurations;
- demonstrate adequate emission controls and combustion efficiency over a range of alternate process and equipment configurations;
- evaluate and/or develop the necessary predictive modeling tools to enable correlation of test results for use in designing larger AFBC utility units with confidence;
- establish a data base for a range of coals and limestones to enable scale-up of AFBC to commercial size; and
- train personnel in the design and operation of AFBC facilities.

C. SUMMARY

Over the past several years significant progress has been made by TVA in developing AFBC technology. The 20-MW AFBC pilot plant is providing a data base for the next generation of utility AFBC units. However, a significant amount of work is still required to optimize boiler designs and support systems. Development of reliable and economical equipment necessary to support the operation of a commercial scale AFBC boiler is an important objective of current test operations. Proper attention in these areas will result in capital and operating costs significantly less than currently estimated. This will contribute to making AFBC technology more competitive than conventional boiler technology.

SHAWNEE POWER PLANT, PADUCAH, KENTUCKY

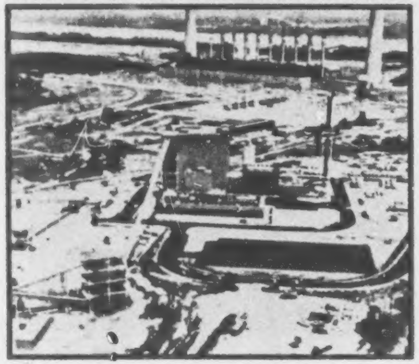
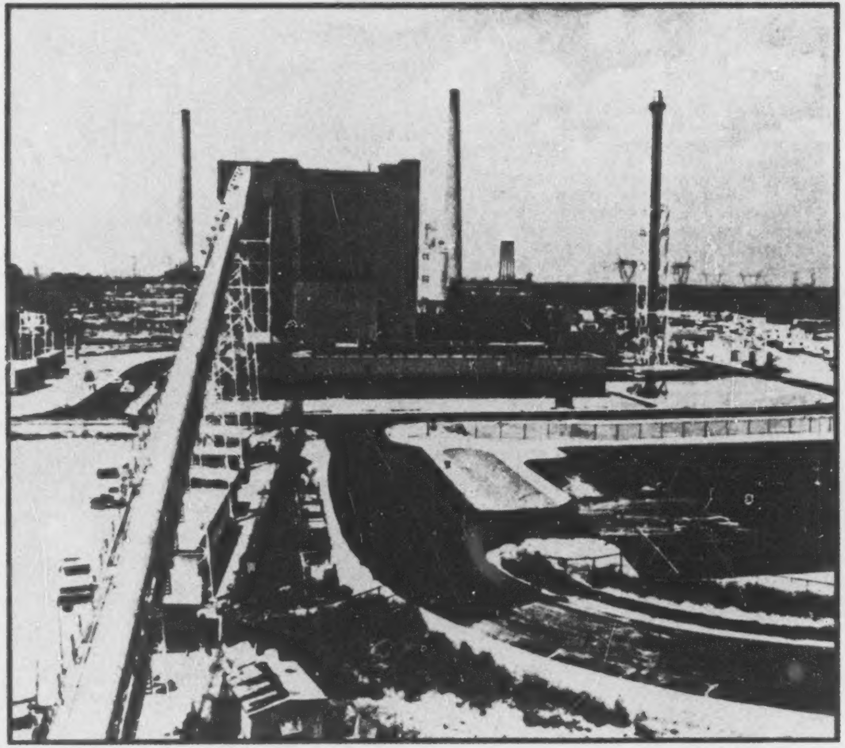


FIGURE 3
TENNESSEE VALLEY AUTHORITY'S 20-MW AFBC PILOT PLANT
PADUCAH, KENTUCKY



IV. Tennessee Valley Authority's Proposed 160-MW AFBC Demonstration Plant

A. INTRODUCTION

TVA proposes to design, construct and operate a 160-MW AFBC plant to demonstrate the commercial viability of AFBC technology. The plant will document the reliability and cost effectiveness of commercial system components thereby lowering the risks that would be likely in going directly from the 20-MW pilot plant to a commercial generating plant of 200-MW or larger.

B. BACKGROUND

TVA contracted in 1979 with three boiler makers to prepare preliminary design studies for the boilers in a 200-MW AFBC demonstration plant and an 800-MW AFBC commercial plant. In parallel with these contracts, TVA's Division of Engineering Design prepared designs and cost estimates for the 200-MW demonstration plant, site, coal and limestone handling facilities, power plant buildings, a turbine/generator, and other balance-of-plant facilities. A performance boiler specification was prepared along with a schedule and cost estimate for a stand-alone 200-MW AFBC demonstration plant.

During the latter part of 1980 and early 1981, TVA reassessed its demonstration project in light of its load forecasts, funding availability, environmental considerations, and the type of facility required to achieve commercialization. It was concluded that TVA would pursue the development of a replacement facility (an addition) rather than a stand-alone plant. An important consideration in this decision was the lower cost — about $200 million compared to an expenditure in excess of $600 million. Other factors were availability of a site at an existing plant and load demand.

Conceptual designs and cost estimates for the demonstration (add-on) plant have recently been completed by TVA. A financial program for the facility is currently being prepared; the facility is scheduled to begin operating in 1989.

C. DESCRIPTION OF THE PROPOSED 160-MW AFBC DEMONSTRATION PLANT

1. Overview

TVA proposes to locate the AFBC demonstration plant at the Shawnee Steam Plant in Paducah, Kentucky. (Figure 4 shows the conceptual layout for the demonstration plant.) To demonstrate the technology, the AFBC boiler will be constructed beside an existing unit at the steam plant as a replacement for the existing boiler. The AFBC boiler will be capable of generating approximately 160-MW and will make maximum use of the existing turbogenerator set and other associated equipment now in use on Shawnee Steam Plant unit 10. The design fuel for the demonstration will be high-sulfur, Illinois Basin coal such as Kentucky number 9. Testing of alternative fuels will be considered on an individual basis during the test program.

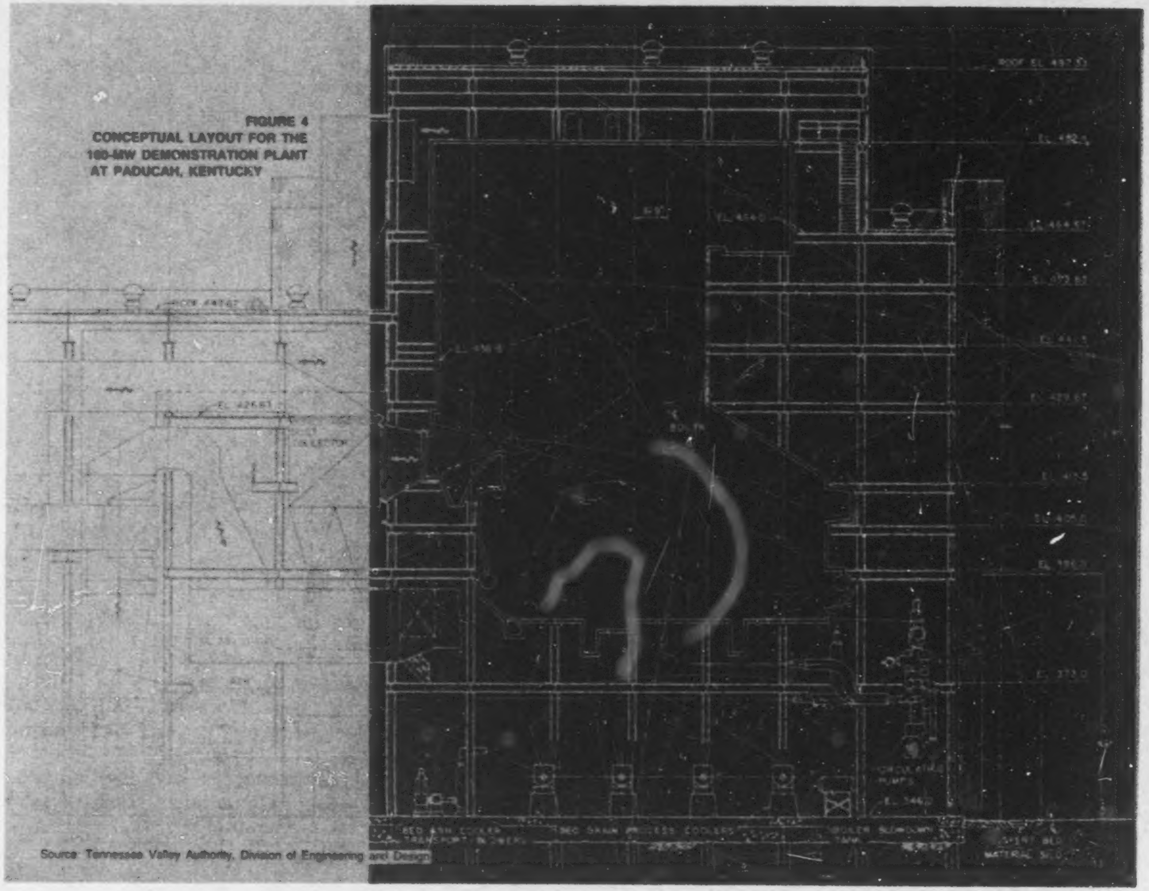

CIRCULATING PUMPS
BED ASH COOLER TRANSPORT BLOWERS
BED DRAIN PROCESS COOLERS
BOILER BLOWDOWN TANK
SPENT BED MATERIAL SILO

FIGURE 4
CONCEPTUAL LAYOUT FOR THE
160-MW DEMONSTRATION PLANT
AT PADUCAH, KENTUCKY

Source: Tennessee Valley Authority, Division of Engineering and Design

2. Boiler Configuration

During the conceptual design phase, a ranch style boiler configuration was selected to be considered for the 160-MW add-on facility. Currently both ranch-style and stacked-bed boiler configurations are being evaluated. No final decision will be made on which to select until further studies and tests have been completed.

3. Fuel Feed and Preparation

The three boiler designs developed during the conceptual design phase used an underbed feed system. Overbed as well as underbed feed systems will be tested at the 20-MW pilot plant. The results of these tests will be taken into account when making the final selection of the feed system to be used in the demonstration plant.

4. Spent Bed Heat Recovery System

The spent bed granular material is discharged from the boiler at approximately 1500°F. This means there is a significant amount of heat in the material which, if recovered, could improve boiler efficiency by as much as 1 percent. Current efforts are directed at developing a fluidized bed cooler which will contain cooling coils submerged within the bed through which feedwater will be routed. The exact details of this concept are still being developed, but it appears to have a significant cost advantage. The design life of such a cooler should also be considerably longer than screw coolers and less expensive to maintain or repair. Other methods of recovering the heat are also being considered.

5. Balance of Plant

The 160-MW AFBC add-on boiler for Shawnee unit 10 will use the existing facilities for coal handling. Separate limestone facilities will have to be added in the yard, because existing limestone facilities do not have sufficient capacity to support an AFBC boiler. Disposal areas serving the Shawnee Steam Plant will be used by the demonstration plant. Fly ash will go to the Shawnee ash pond. Spent bed material will be collected dry and disposed of separately.

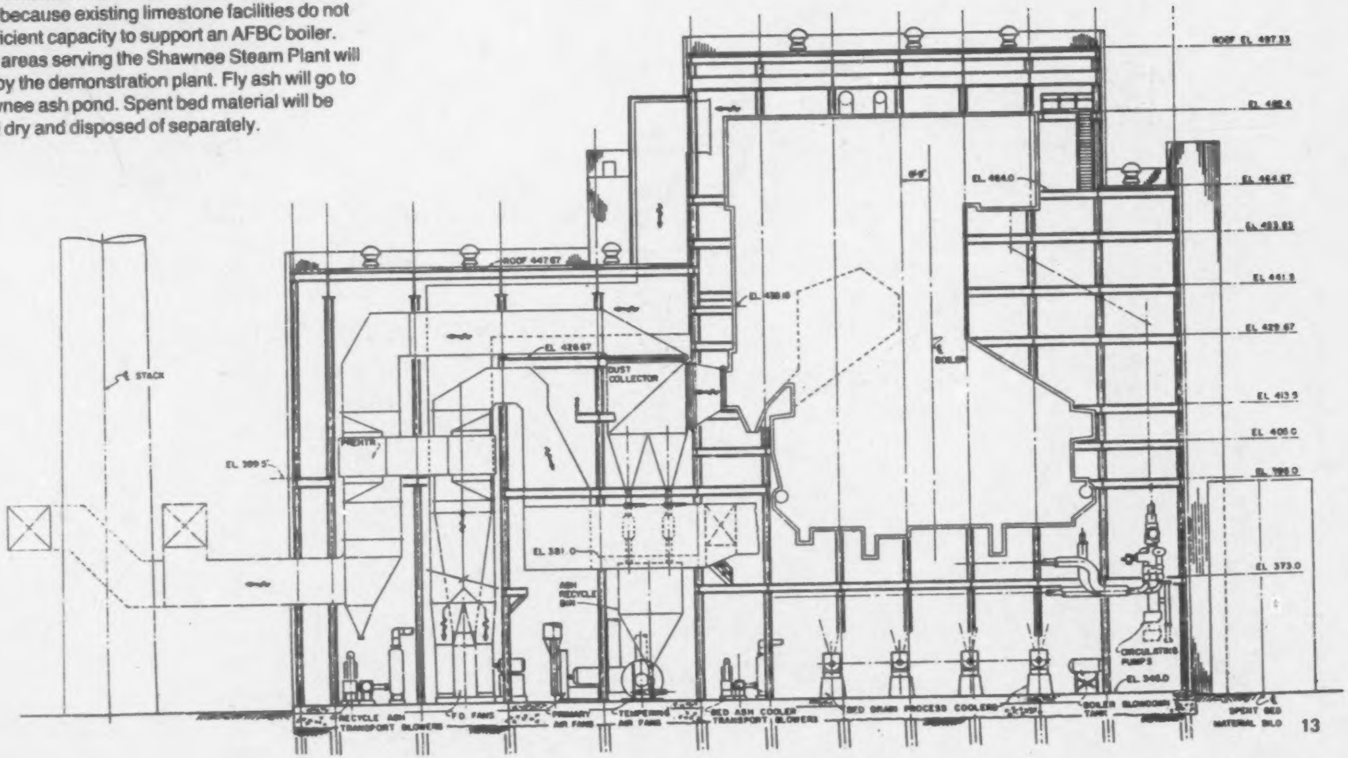
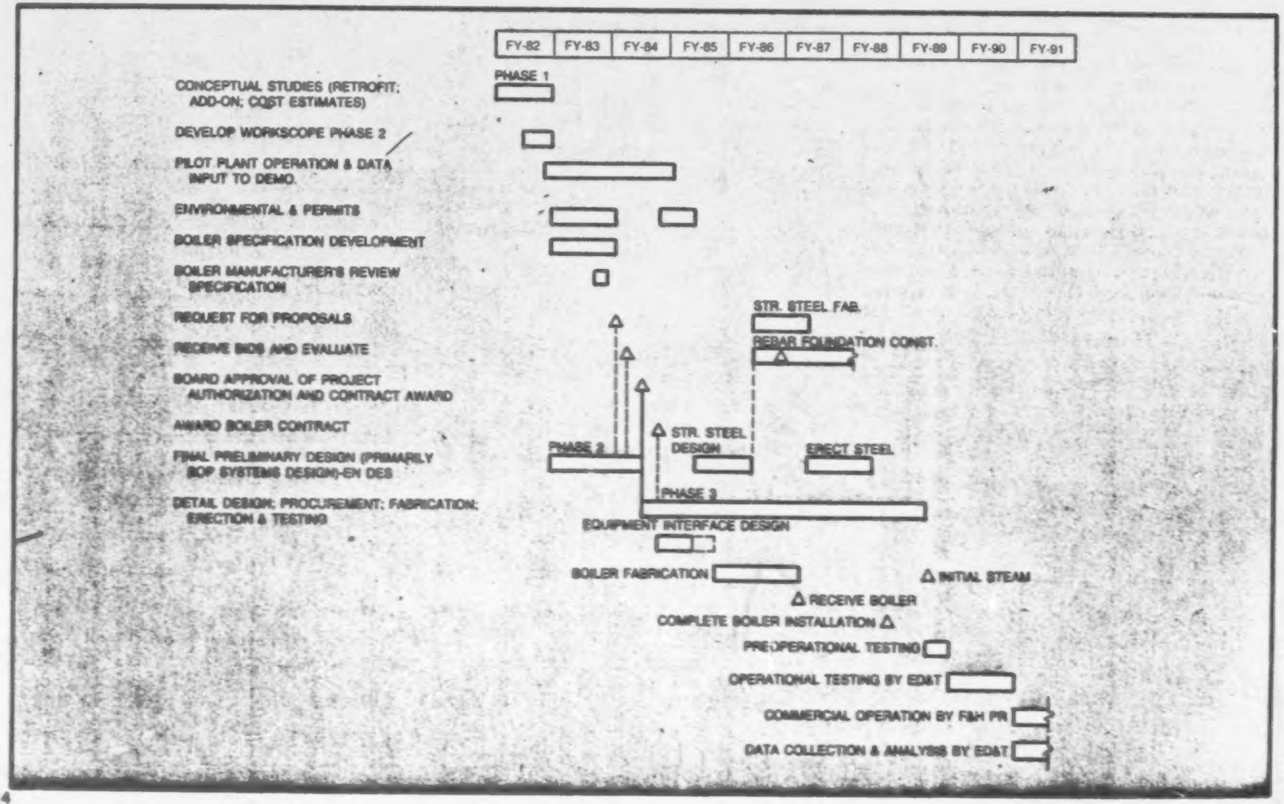

ROOF EL 497.33
EL 482.4
EL 464.67
EL 453.83
EL 441.5
EL 429.67
EL 413.5
EL 405.0
EL 395.0
EL 373.0
EL 464.0
EL 438.0
BOILER
ROOF 447.67
EL 426.67
DUST COLLECTOR
STACK
PREHTR
EL 399.5
EL 381.0
ASH RECYCLE BIN
CIRCULATING PUMPS
EL 346.0
RECYCLE ASH TRANSPORT BLOWERS
F D FANS
PRIMARY AIR FANS
TEMPERING AIR FANS
BED ASH COOLER TRANSPORT BLOWERS
BED DRAIN PROCESS COOLERS
BOILER BLOWDOWN TANK
SPENT BED MATERIAL SILO

FIGURE 5
PROPOSED TVA AFBC DEMONSTRATION SCHEDULE



D. CURRENT AND FUTURE ACTIVITIES

Conceptual design activities have been completed and preliminary design work is presently underway to develop the add-on approach. The results of these initial activities were the identification of the balance-of-plant modifications required, development of conceptual designs for the modifications and for the boiler, and a schedule and cost estimate for the facility. For these studies, a ranch-style boiler was chosen as being representative of an AFBC boiler. However, no decision has been made whether or not future design activities will be based on a ranch or stacked-bed boiler configuration.

The present schedule allows a reasonable amount of time to evaluate the pilot plant test results prior to "freezing" boiler designs. On this schedule, detailed design and erection of the demonstration facility will begin during the summer of 1984 and will be completed by mid-1989 (see Figure 5). Operational testing with power to the grid will begin in mid-1989. This timetable provides TVA or other utilities the option of selecting an AFBC coal-fired plant in the late 1980s or early 1990s to provide added or replacement power to the grid in the mid-to-late-1990s.

The cost advantages of AFBC compared to conventional PC boilers with FGD are expected to be important incentives to the use of this new technology by the utility industry. Studies by TVA, EPRI and others indicate that capital costs of an AFBC boiler for a 200-MW plant will be 10 to 15 percent less than for a conventional PC boiler with FGD. Also, based on studies, operating costs for AFBC boilers are expected to be 10 to 20 percent lower than for conventional PC boilers with FGD.

Preliminary cost estimates for the 160-MW AFBC boiler at Shawnee are $200 million (in current dollars). In addition to its own funding, TVA is seeking participation from the United States and foreign governments, utilities, institutions and companies to finance the development and testing of this AFBC boiler. Development of the 160-MW AFBC boiler will be an important step toward the commercial use of AFBC technology in the utility industry.

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This November 1979 article from Power Engineering reviews the status of atmospheric fludized bed combustion to that date. Efforts by leading manufacturers of AFBC systems are described and examples of units in operation provided. For the most part, these are small industrial units only.

Fluidized bed combustion systems:

FBC presents a way to burn coal with minimal SO_2 and NO_x emissions. Development work is leading into demonstration units by a number of manufacturers

By I. M. BERMAN, Senior Editor

Combustion in a coal-fired boiler usually takes place at 2500 to 3000 F, and with excess air as needed to insure complete combustion of all the carbon. This combustion usually produces the following undesirable effects:

1. Sulfur in the coal combines with oxygen in the combustion air to produce gaseous sulfur dioxide (SO_2).

2. Nitrogen in the coal and in the air combines with oxygen in the air to produce gaseous oxides of nitrogen (NO_x).

3. Particulates (ash and other noncombustible constituents) are carried to the stack with the flue gases.

4. Heavy metals, such as cadmium, are liberated.

5. A considerable amount of hard ash particles, usually with sharp edges, is formed.

At present, the scrubber, wet or dry, is the best available technology for removing SO_2 from flue gases. Scrubbers are quite expensive, with some estimates for the complete flue gas desulfurization system running as high as 25% of that of the basic plant. In addition, they require a large land area and constitute a major system that must be maintained and repaired. A breakdown in the scrubber may mean a shutdown of the entire plant.

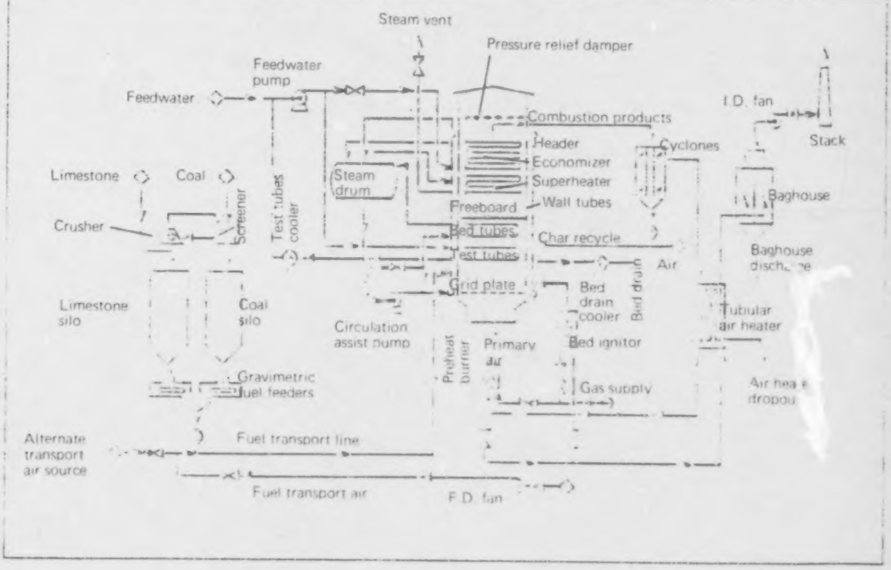
The equipment for removing NO_x from flue gases is only lately emerging from the laboratory, but for new steam generators with new burner designs, it is possible to meet the current NO_x standard of 0.6 lb/million Btu without auxiliary equipment.

Fluidized-bed combustion

Control of SO_2 and NO_x emissions would be simplified if there was some way to entrap the sulfur during combustion, and if at the same time the coal could be burned effectively and efficiently at a temperature below 2000 F so fixation of atmospheric nitrogen would not occur. Fluidized-bed combustion (FBC) offers the capability of doing this. Its use in power generation is a recent development, with demonstration units now in design and in the early stages of construction.

A fluidized bed is a layer of small, noncombustible particles resting on a perforated plate. Air is forced up through the perforations to give this bed a sort of fluid motion. Fuel is added to the bed and burned in the bed. The material, size of the perforations in the bed plate, fluidizing air velocity, and ambient temperature are all controllable for the exact effect desired. For example, the combustion of coal in a fluidized bed uses limestone with a high concentration of lime (calcium oxide).

Figure 1. Flow diagram of Combustion Engineering's sub-scale fluidized bed test unit. *(Diagram courtesy of Combustion Engineering Inc.)*


Steam vent
Pressure relief damper
Feedwater pump
Feedwater
I.D. fan
Stack
Combustion products
Header
Economizer
Superheater
Cyclones
Limestone
Coal
Steam drum
Baghouse
Crusher
Screener
Test tubes cooler
Freeboard
Wall tubes
Bed tubes
Char recycle
Test tubes
Air
Baghouse disch
Grid plate
Bed drain cooler
Bed drain
Limestone silo
Coal silo
Tubular air heater
Circulation assist pump
Preheat burner
Primary air
Bed ignitor
Gravimetric fuel feeders
Gas supply
Air hea
dropou
Alternate transport air source
Fuel transport line
Fuel transport air
F.D. fan

Progress and outlook

FBC provides the following advantages:

1. The coal is burned in the bed in intimate contact with the limestone. Virtually all the sulfur is captured by the lime, and very little SO_2 is discharged with the flue gases.

2. Combustion normally takes place between 1500 F and 1750 F—hundreds of degrees below the point where atmospheric NO_x formation becomes troublesome.

3. Depending on the type of coal used and the auxiliary equipment employed, particulate carryover may be less than in a conventional coal-fired boiler.

4. Carbon burnup may be as high as 99.9%.

5. Virtually any type of combustible material may be burned in an FBC by properly adjusting factors such as type and particle size of bed material, fluidizing velocity, feed methods, and rate of feed. Sulfur content of fuel can be over 3.5%, and the process will still limit SO_2 discharged to below current EPA limits without the use of a scrubber.

Lignite, compared to normal coal, has a very high proportion of fines. ("Fines" are the smaller particles, compared to the desired size range.) Care must be taken when introducing lignite into the combustion zone (bed) of an FBC, but as long as the lignite is burned in the bed and not above it (in the "freeboard" volume), satisfactory heat release and flue gas control can be effected.

The fluidized bed combustor is a rather simple device (see Figure 1). Before the air is turned on, the bed is at rest with the top of the bed at the "slump" level (below "test tubes"). After the air is supplied, the bed top rises. The ratio between heights of the fluidized level and the slump level is usually between 2:1 and 4:1. The bed is heated by auxiliary burners to about 1100 F. Coal is added directly into the bed and undergoes ignition at this point. If a spray stoker or similar device is used, coal is now added to the bed, and ignition occurs. Burning of the coal raises the combustion temperature to the normal operating level above 1500 F.

Steam is generated primarily in water tubes that are immersed in the bed after fluidization. Depending on the design, the waterwalls (vertical water-filled tubes in the side walls of the combustor) either generate steam or act as multiple downcomers to provide water for the in-bed tubes.

Superheating of the steam normally takes place in a group of tubes located in the freeboard volume, and some designs place economizer (feedwater preheater) loops above the superheater. Figure 1 is simplified in that it does not show steam-side components such as drums, downcomers, and piping connections.

There are two general classifications of FBCs: atmospheric and pressurized. As the names imply, the combustor in an atmospheric FBC operates at or very near ambient (atmospheric) pressure, while the pressurized FBC may be pressurized to between three and 10 atmospheres. This latter arrangement calls for additional auxiliary equipment to maintain pressure during operation.

Figure 1 shows, in schematic form, a typical relationship between the combustor and many of the auxiliaries. As the process is now structured by all manufacturers, coal either of the correct size or larger than required, and pre-sized limestone, are delivered to the FBC installation. The coal is crushed to the correct size, and the coal and limestone are stored in silos. As coal is needed, it is withdrawn via the gravimetric coal feeder and blown or thrown into the bed. Coal feed is determined by the heat rate required. Limestone is fed into the combustor in the same manner. The need for additional limestone is determined by checking the level of discharged SO_2. If the level begins to rise, limestone is added.

Combustion air ("primary air" in Figure 1) is provided to a plenum under the grid plate. It may be heated by a separate preheat burner or warmed by exhaust gases in a heat exchanger, or both preheating methods may be used.

Flue gases pass through one or more centrifugal particle removers (usually cyclone separators). The particulate discharge from these cyclones may contain up to 15% unburned fuel, which is recycled by reinjecting it into the combustor. Cyclone separation and reinjection of the residue will result in a carbon burnup efficiency of better than 99.5%.

Not all particulates are recycled. Those that pass through the separation process are filtered out in a baghouse. The discharges from baghouses and other auxiliaries in the flue gas cleanup system are removed and can be used for landfill. However, the amount of elutriated material (carried over in the flue gas) is often less than in a conventional unit.

The combustion bed is also drained, usually when additions of limestone and reinjection of cyclone discharges cause bed level to increase beyond the desired point. The bed material drawn off consists almost entirely of spent and unspent limestone, and ash. It is also quite suitable for landfill.

All other systems shown in Figure 1 are conventional.

Several advantages—those having the greatest impact in view of environmental restrictions—have been mentioned above. Another advantage for the FBC is the physical difference in the ash. A dry, powdery form is produced in an FBC, while combustion above 2000 F may result in slagging, "clinker" ash that is difficult to handle. Also, there is a reduced discharge of heavy metals and alkali salts with the flue gas.

The range of 1500-1750 F has proved most efficacious in minimizing SO_2 and NO_x discharge. The pressurized FBC apparently operates best between 1700 F and 1750 F, while a temperature range about 150 F lower is preferable for the atmospheric process. This temperature requirement is not expected to cause any great difficulty in designing and operating an FBC.

Combustion in a fluidized bed has been used for several decades. Originally, its biggest use was in the chemical industry where it was used for such processes as coal gasification in the 1920s. No additional use was made of the technique until World War II, when it led to catalytic cracking of petroleum to produce high-octane aviation gasoline. Additional processes have developed since 1945, and there are a number of methods of producing combustible gases and liquids in which a fluidized bed is used.

During the same postwar period, experiments began in the generation of steam using hard-to-burn materials in a fluidized bed. It soon became obvious that anything combustible would burn in an FBC and, with bed temperatures of 1500 F to 1750 F, no special alloy is required if tubes are actually located in the hot bed.

About five years ago, small FBC steam generators began to be developed, all with steam-generating tubing in the bed. The original horizontal orientation of bed tubes has all but disappeared, as the tubes tended to interfere with free movement of the bed material during fluidization. Also, the deeper the bed, the greater the need to orient the in-bed tubes more toward the vertical.

FBC SYSTEMS

A nice balance must be struck by the FBC designer and operator among three factors: gas (fluidizing) velocity, particle size,and bed depth. The use of steam-generation tubes in the fluidized bed results in a considerably higher heat-transfer rate then would occur if the tubes were heated only by hot gases. According to studies by Babcock Contractors, Inc. (the construction arm of Babcock & Wilcox, Ltd., of England), the heat release per unit area of a fluidized bed is a linear function of the velocity of gases leaving the bed. That is, the higher the velocity, the greater the heat release. Care must be taken, however, for as the velocity is increased, the size of the bed particles (coal and limestone) must be increased so that the particles will not become entrained in the gas but will remain in the bed. Another factor is that the rate of heat transfer decreases as particle size increases.

Also, as fluidizing velocity increases, the bed must usually be made deeper to provide both adequate residence time for small particles and room for sufficient heat transfer surface area. The optimum system balances all these factors.

A number of companies have been devoting considerable effort to developing commercial FBC systems. As a rule, a company starts off with a small unit, with a combustion area of only a few square feet, and as the concept of FBC proves out, larger units are planned and built. Several companies' efforts will be examined in detail. Selection of these companies should not be construed as indicating they are the only ones working on FBC; the development effort is worldwide, as evidenced by the selection of examples to follow.

Babcock Contractors, Inc.

Babcock has been operating a 45,000 lb/hr fluidized bed boiler at Renfrew, Scotland, since May, 1975. The boiler was originally designed to produce 40,000 lb/hr of steam at 400 psig and 560 F. The firebox had a spreader grate, which was replaced with a bed about 9 ft x 9 ft in size. Horizontal tubes were added in the bed and connected via circulating pumps to the original steam generating circuit. Air is supplied under the perforated grid; coal and limestone, as well as recirculated unburned fuel particles, are injected into the bed under pressure. The colored portions of Figure 2 show the boiler components changed or added to allow fluidized bed operation.

The operation of the Renfrew FBC has been so successful that Babcock, in conjunction with Riley Stoker, has been chosen by the Ohio Governor's Coal Use Committee to provide steam at several state institutions and industrial boiler plants. The first of these units will be an FBC retrofit into a 60,000 lb/hr conventional boiler at the Central Ohio Psychiatric Hospital in Columbus.

Babcock and Riley plan to provide units with steam capacities as high as 500,000 lb/hr for use by refineries that do not wish to burn their own product for plant use, as well as by municipalities for trash incineration, and by institutions. A boiler of that size could also be used by electric utilities. The rule of thumb in boiler rate vs generator output is that 8000-10,000 lb/hr of steam is equivalent to 1 MWe. 500,000 lb/hr could produce as much as 60 MWe.

Combustion Engineering, Inc.

Combustion Engineering, Inc. (C-E) efforts began in the early 1950s, when the company spent one year on a gas- and oil-fired FBC development program as an alternative to conventional firing systems. The technology was not sufficiently advanced, and there was no economic justification, so the program was dropped in 1953. With the sudden upsurge in environmental considerations, however, the firm began developing a coal-fired combustion system that would meet the SO_2 and NO_x emission requirements.

C-E first built two- and three-dimension cold-flow models of transparent plastic to obtain information on air distributor plate hole sizes and patterns and bubble characteristics. These models provided information on operating a fluidized bed with one section active and an adjacent one slumped, and fluidizing one section from an adjacent one.

Armed with this knowledge, C-E designed and built a sub-scale unit (SSU) (see Figure 3) to simulate one modular unit in a larger plant. While the bed has an area of less than five square feet, it uses all the components of a larger system.

The SSU can provide 2300 lb/hr of superheated steam at 365 psi and 560 F. It uses 300 lb/hr of coal and 100 lb/hr of limestone.

It should be noted that the amount of coal is only a small percentage of bed material. Since combustion tends to be almost complete while in intimate contact with the hot limestone, coal rarely occupies more than 2% of the bed volume, regardless of the process used and the pressure in the combustor.

There are four steam generation tubes in the bed volume. These tubes incline about 10 deg above the horizontal to ensure the natural circulation, and they are connected to the waterwall tubes, where evaporation also occurs.

Coal particles are nominally ¼ in. long, and limestone ⅛ in. long. The mole ratio of calcium in the limestone to sulfur in the coal is a nominal 3.0, and the fluidizing air velocity is 7 fps. C-E has found that the angled bed tubes tend to reduce bed splashing into the freeboard volume.

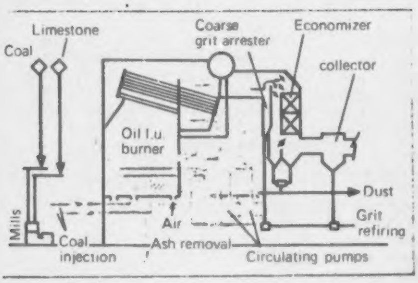
C-E has completed over 2000 hours of testing in its SSU, most of it with two Illinois No. 6 high-sulfur coals. The results have been most encouraging, and have led to a contract to build a 50,000 lb/hr demonstration plant at the Great Lakes Naval Training Station in northeastern Illinois. C-E is sharing the cost of this program with the U.S. Navy and the U.S. Department of Energy. The demonstration plant boiler will have a bed area 8 ft wide x 17½ ft long. Bed depth when fluidized will be 3 ft.

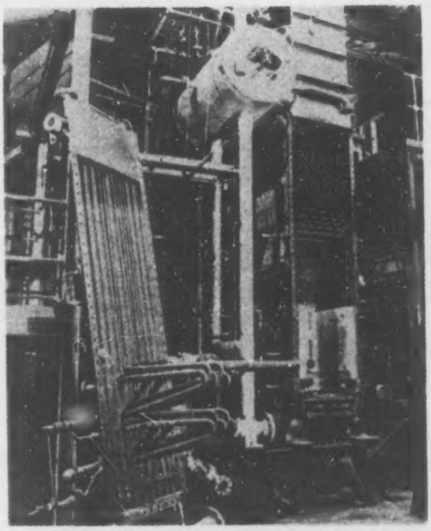
The boiler is shown in Figure 4. Only one bank of steam generating tubes is shown; the nearer bank has been removed for clarity. The short horizontal-U tubes are in the bed volume during fluidization. The combustion gas flows up, then turns downward over the waterwall tubes, generating more steam. Gas exit is outboard of the U sections. Superheating is performed by a small number of long, U-formed tubes supported by piping headers outside the boiler rear wall and by attachments (not shown) to the bed tubes.

The superheater does not extend the entire length of the boiler, for during start-up only a small amount of steam is produced, and the superheater tubes would not be adequately cooled. The bed units nearest the front of the boiler (no superheater tubes above) will always be lit off first when starting from "cold iron." As steam begins to flow, fluidization of other bed sections will begin.

Ignition air will be heated to 1400 F and injected directly into the bed by means of nozzles in the bed. These ignition nozzles are adjacent to the coal feed nozzles, so that the heated air will cause the nearby coal to ignite. Heating only restricted areas to get bed combustion saves fuel that would normally be used by over-bed heaters in bringing a complete zone to a self-sustained coal combustion temperature.

When the bed is slumped, it presents a greater resistance to air flow than it does after fluidization. In order to avoid a high pressure drop across the bed, an under-grid air bypass system is employed. This system effectively provides more than twice the grid plate open area, maintaining the pressure drop across the bed at a value consistent with normal operating conditions. As combustion begins, hot gases are produced and the particles' motion is enhanced. The bypass system can then be cut off. This arrangement allows the use of smaller forced draft fans than might otherwise be required.





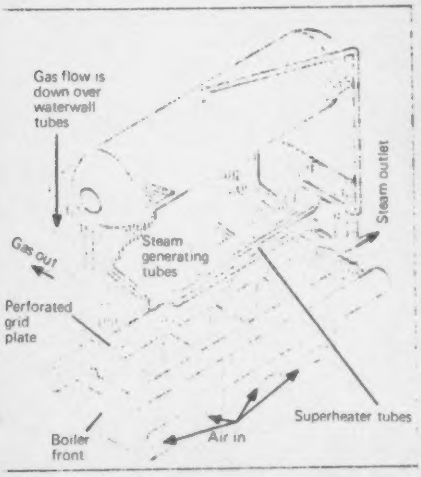


Figure 2. *(upper left)* 40,000 lb/hr Babcock steam boiler in Renfrew, Scotland, modified for an FBC experiment. The original spreader grate was removed and replaced with a fluidized bed. Boiler components modified to accommodate the FBC are shown in color. *(Diagram courtesy Riley Stoker Corp. and Babcock Contractors, Inc.)*

Figure 3. *(above)* Combustion Engineering's sub-scale fluidized bed test unit. The front of the unit has been removed and set to the left. The horizontal V shaped tubes are submerged in the bed when it is fluidized. *(Photograph courtesy of Combustion Engineering, Inc.)*

Figure 4. *(left)* General arrangement of fluidized bed demonstration boiler. A unit of this type will be installed at Great Lakes Naval Training Station in Illinois.

The boiler has eight zones, each 4 ft wide x 4½ ft deep. Output is regulated most easily by changing the number of zones in operation. In addition, a temperature variation from 1400 to 1700 F permits a change of ± 10% of nominal output per zone. Ignition of any zone after the first is by heat transfer from an adjacent zone.

The demonstration unit will provide 50,000 lb/hr of steam at 365 psi and 560 F to the main header at the existing plant. 3.5% sulfur Illinois coal will be used, and a 4:1 calcium-sulfur mole ratio is planned.

New York State funded a study on retrofitting Consolidated Edison's Arthur Kill plant—a C-E boiler originally designed to burn coal but used only with oil—with atmospheric FBC units. C-E's concept was a series of FBC banks stacked four high around the periphery of the tall combustion volume. Hot gases would flow up a central stack. Estimated cost for this retrofit would run $60 to $70/kW more than using a pulverized coal combustor and a scrubber.

C-E's preferred basic design has all the beds on one level (as opposed to the Arthur Kill study). The company feels that feeding fuel from the side, breaking up beds into small modules, convenient regulation of air flow, and ease of maintenance more than justifies the extra land required.

Another aspect of the future was presented when the Tennessee Valley Authority (TVA) requested C-E—as well as two other FBC manufacturers—to provide a conceptual overview of a 200-MW FBC demonstration plant to be completed in the late 1980s or early 1990s. This plant, including the studies, would be a multi-billion-dollar undertaking, and if successful could lead eventually to FBC boilers with capacities up to 1000 MWe.

Foster Wheeler Corp.
Foster Wheeler Corp. (FW) has four demonstration units in various stages of design or completion:

1. Monongahela Power Co. in Rivesville, W. Va.—300,000 lb/hr of steam added to the main steam manifold of the plant (completed).
2. Georgetown University, Washington, D.C.—100,000 lb/hr of steam to be added to the steam manifold of the heating and cooling plant (under test).
3. DOE Morgantown Energy Technology Center, Morgantown, W. Va.—60,000 lb/hr nominal (90,000 lb/hr peak) component test and integration unit, with steam output used for heating (to be started in 1981).
4. City of Wilkes-Barre, Pa.—100,000 lb/hr of steam to be used in the city heating system (to be started in 1981).

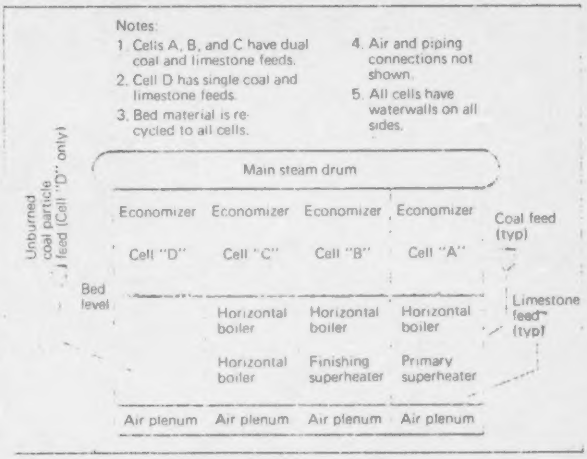
FW's first major effort was the first development of its kind in this country to "go commercial," as far as electric power generation is concerned. The development project at Rivesville began in late 1968 and became operational in 1976. Because the technology was so new, it was felt that using the FBC as an adjunct to the main steam system would give the manufacturer and the utility experience in FBC operation but would not jeopardize the generation of electricity. The Rivesville installation is shown diagrammatically in Figure 5.

FW's Georgetown University plant, as of this writing, is nearing completion of startup testing. This unit will provide 100,000 lb/hr of steam at 625 psi and temperatures to match the main heating plant steam manifold, using coal with 3.3% sulfur and 1 in. x ¼ in. in size.

At full load coal will be consumed at the rate of 9500 lb/hr, while the limestone feed will provide 3100 lb/hr. The thermal efficiency will be about 83.5%, but can be raised with air preheating. A cross-section view of the Georgetown plant is shown in Figure 6.

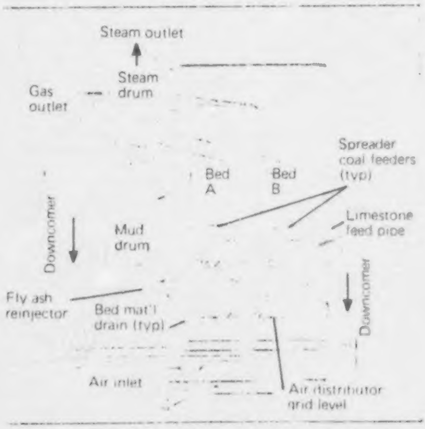
The grid plate is made in sections and is flanged together. Holes are drilled in a predetermined pattern, bolts are inserted and tack welded in place, and each bolt is drilled through longitudinally to form a three-inch nozzle through which fluidizing/combustion air passes. This will allow an insulating layer of limestone and ash nearly 3 in. deep to remain relatively at rest on the grid plate. The Rivesville plant used a 16-gage plate with many small holes drilled on ½ in. centers, and thermal stress cracks developed between the holes.

The Georgetown unit uses the FW spreader stoker, and to keep fines from burning in the freeboard the fuel is dampened so that all particles fall into the bed for combustion. Limestone, slightly more abrasive than coal, falls via a drop pipe into the bed, where it is distributed by the fluidizing motion.



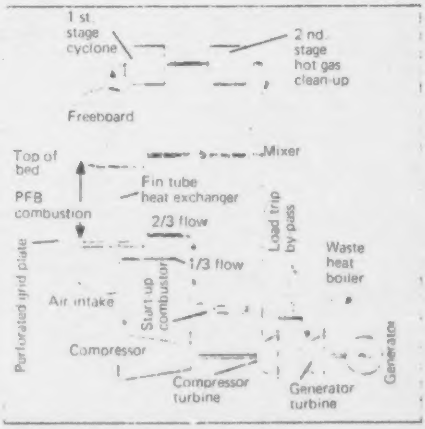
Notes:
1. Cells A, B, and C have dual coal and limestone feeds.
2. Cell D has single coal and limestone feeds.
3. Bed material is recycled to all cells.
4. Air and piping connections not shown.
5. All cells have waterwalls on all sides.
Unburned coal particle feed (Cell "D" only)
Main steam drum
Economizer
Economizer
Economizer
Economizer
Coal feed (typ)
Cell "D"
Cell "C"
Cell "B"
Cell "A"
Bed level
Horizontal boiler
Horizontal boiler
Horizontal boiler
Limestone feed (typ)
Horizontal boiler
Finishing superheater
Primary superheater
Air plenum
Air plenum
Air plenum
Air plenum

Figure 5. *(right)* Layout of the Rivesville FBC demonstration plant. The Foster Wheeler Energy Corp. supplied this 300,000 lb/hr boiler to Monongahela Power, and steam was first generated in late 1976.



Steam outlet
Gas outlet
Steam drum
Spreader coal feeders (typ)
Bed A
Bed B
Downcomer
Mud drum
Limestone feed pipe
Fly ash reinjector
Bed mat'l drain (typ)
Downcomer
Air inlet
Air distributor grid level

Figure 6. *(below left)* Steam generator arrangement of the Foster Wheeler/Georgetown University FBC boiler. *(Diagram courtesy of Foster Wheeler Energy Corp.)*



1 st stage cyclone
2 nd stage hot gas clean-up
Freeboard
Top of bed
Mixer
PFB combustion
Fin tube heat exchanger
2/3 flow
Load trip by pass
Perforated grid plate
1/3 flow
Waste heat boiler
Air intake
Start up combustor
Compressor
Compressor turbine
Generator turbine
Generator

Figure 7. *(below right)* Pressurized fluidized bed heated air flow diagram. Combustion air can be diverted at the mixer so that only heat exchanger air operates the compressor turbine. *(Diagram courtesy of Curtiss-Wright Corp.)*

FW does not use cyclone separators for the flue gas but prefers 180-deg turns by means of baffle plates. At the first turn, unburned particles drop out and are returned to the combustor. Eighty to 85% combustion efficiency is achieved on the first pass, and char recycle improves that to greater than 95%.

Georgetown uses about 2% to 3% carbon by volume in the bed, and the combustor can provide a 4:1 turndown ratio by adjustment of fuel and air rates, bed temperature, and bed depth, and control of the number of zones in use.

An igniter will start up one zone at the end of the bed, and as soon as the operating temperature is attained, all zones will be fluidized and combustion will propagate the length of the bed. FW indicates that the entire system can be brought up to full operation in about 1½ hr, with the major concern being that of carefully monitoring steam drum deflection.

The FBC walls consist almost entirely of water tubes in which steam is generated. The earlier developments used refractory-lined combustion volumes, which were particularly useful in incinerating garbage and sewage sludge, both containing high amounts of moisture, and in burning off pollutants that would have been difficult to dispose of otherwise.

The Department of Energy (DOE) Morgantown Energy Technology Center unit will generate 60,000 lb/hr of steam, with a peak capacity of 90,000 lb/hr, in a three-tiered arrangement. The hot gases will pass up a common flue, and the unit will provide a fully instrumented test combustor for FBC design optimization.

The Wilkes-Barre unit will use anthracite culm as fuel. This is a residue from the burning of anthracite, 60 to 70% inert, and it appears as gray, shale-like gravel. There is heating value left in this culm, but to date no process except FBC will make it burnable. A key to making this new-found source of fuel combustible is tumbling and agitation in the fluidized bed that continually wears away the outer covering, exposing combustible material.

FW indicated that the heat transfer rate in FBC is about 50 Btu/hr-ft²-deg F in the steam generation tubes immersed in the bed; boiler bank tubes, heated only by hot flue gases, transfer only 10 Btu/hr-ft²-deg F to the water inside. In some designs, there may be no need for any boiler bank steam generating or superheating tubes at all.

While immersing the tubes in the bed would seem to subject them to rather rugged conditions, erosion is minimal compared to boiler tubes. The constant tumbling and rubbing of the coal and limestone rapidly wears away the sharp edges of each particle. Motion is slower—less than 8 fps for FBC, compared to about 50 fps in a conventional boiler—and because the temperature is considerably lower than the slagging temperature of the ash, no hard particles are formed that can impact the piping and cause wear.

Instead, the ash is soft and powdery and acts almost as a lubricant on the outsides of the tubes. The same low temperature means that the in-bed tubes would not be rapidly affected by heat.

FW is investigating a 200-MW atmospheric FBC system, with steam temperature of 1005 F. This fluidized bed unit either could be connected to the steam manifold of any utility plant or could be used as its main heat source.

FW-supplied data shows no clear superiority for FBC in such factors as capital investment, particle emissions, requirement for excess air, thermal efficiency, direct operating manpower, cost of chemicals for SO_2 capture, or time for construction. In fact, in some of these considerations, an FBC unit may actually be slightly inferior to the best designs in pulverized coal systems. Where FBC is superior, however, is in sulfur removal without the need for a scrubber, NO_x emission control without expensive flue gas cleanup equipment, broad tolerance to changes in coal quality, excellent turndown ratio, and smaller full load auxiliary power consumption compared to coal systems that use scrubbers.

Curtiss-Wright Corp.
Curtiss-Wright Power Systems (C-W) is progressing down a parallel path: that of the pressurized fluidized bed combustor (PFBC). C-W is designing a complete cogeneration system, calling on company experience in gas turbine generating technology to marry a gas turbine-alternator to a PFBC and market the combination as a package.

Curtiss-Wright does not consider itself strictly as a pressurized FBC designer and supplier. One design, almost complete, will provide an atmospheric FBC in Shamokin, Pa. to burn anthracite culm to provide steam for a paper plant. This culm is quite similar to that used in the FW system, having a heating value of 3000 to 5000 Btu/lb and an ash content of 65 to 70%.

The product line that C-W is pursuing most actively in this field, however, is the PFBC. C-W's PFBC is an air heater device, as compared with those FBCs that evaporate water into steam in the waterwall and bed tubes. The outlet of the combustor must be pressurized in order to drive the gas turbine. The bed, as in the other systems, consists of dolomite (a form of limestone) and ash. The flow diagram is shown in Figure 7.

About one-third of the air supply actually passes through the grid plate to fluidize the bed and support combustion. The balance of the air passes through an in-bed heat exchanger consisting of a series of vertical tubes finned inside and out, and eventually joins the air heated directly in the bed. The air that has been used for bed combustion passes through cyclone particle separators and hot-gas cleanup equipment, is combined with the hot gases from the heat exchanger, and drives the compressor that supplies the fluidizing air. The hot gas flow then operates a free gas turbine (no mechanical connection to the PFBC air compressor) which is on a common shaft with an alternator. The remainder of the heat in the gas stream powers a waste heat boiler which in turn operates a steam turbine-generator set. Heat flow rates and equipment ratings have been chosen so that the gas turbine will provide 60% of the system's electrical output, and the waste heat boiler will supply 40%.

Some of the advantages of this method, according to C-W, are that two-thirds of the heated air and gases, while maintaining full heat value, have no particulates, SO_2, or NO_x from the combustion process. In addition, part or all of the fluidizing air can be diverted, i.e., not mixed with the heat exchanger air, in order to provide heated air to the turbine that is clean and free from harmful chemicals. In this manner, the turbine, which is sensitive to even small amounts of gases such as SO_2, and to particulates, will last considerably longer. The bed can still be run at a constant 1650 F for best sulfur capture. In addition, there is a reduction in the total flow needed for fluidization and combustion, and when the flow is reduced, the amount of coal needed to maintain combustion is less.

C-W claims the vertical tube orientation produces no tube erosion. The vertical tubes also permit a deeper bed (16 ft) to be used, allowing greater residence time of coal particles in the bed for greater sulfur capture. A velocity of 3 fps provides sufficient turbulence without risking a significant amount of bed material carryover. Incipient fluidization occurs at an air velocity of 1½ fps, according to C-W.

C-W has built and operated a very small

PFBC which they call a "technology" model. The fluidized-bed unit was one supplied by C-W's Dorr-Oliver subsidiary (see Figure 8). A small turbine was operated with the air heated in the technology model.

Testing of this initial model has been completed, and the results have been factored into the design of the pilot plant. This design is also completed, and construction will begin in the very near future. A 1981 completion is planned.

A two-year operational period is scheduled. The main objectives will be to evaluate the design with respect to reliability, cost durability, and performance. At that point, extrapolations of the data can be made to permit the detailed design of a full-scale commercial plant.

Commercial plant design goals, as C-W sees them, are to burn coal with a sulfur content greater than 3%; meet all EPA emission requirements; and have conventional-unit capabilities and fuel costs. Plans are to design this installation and use it for intermittent duty—operation will be during the regular five-day work week. Bed heat loss is projected to be only 150 F in a 24-hour period, so, assuming a normal weekend, the bed will be near 1100 F on Monday morning, facilitating rapid coal ignition and restoration of output. A cold iron start, on the other hand, will require about eight hours to full output, for this plant will have a refractory-lined combustor, as opposed to the waterwall construction common to atmospheric FBC steam plants.

The vertical heat exchanger tubes are perhaps the most interesting part of C-W's design. The tubes are connected to the grid plate at the bottom of the bed. Air supplied to these tubes flows through a different plenum than does the fluidizing/combustion air. These are counterflow tubes, with an inner tube not in contact with the inner fins of the outer tubes (see Figure 9). Air from the plenum flows up the annulus, being heated by contact both with the outer wall and the inner fins. A cover at the top of the tubes prevents the heated air from escaping and directs it down the inner tube, so that it will lose almost none of the heat it acquired while rising. To insure this minimum heat loss, the inner tube is itself a double-wall tube, with insulation between the two walls. The heated air flows out from a manifold at the bottom of the vessel.

C-W estimates that total generating costs could be about 10% lower than those of a pulverized-coal conventional unit with a scrubber, and 30% lower than that of an

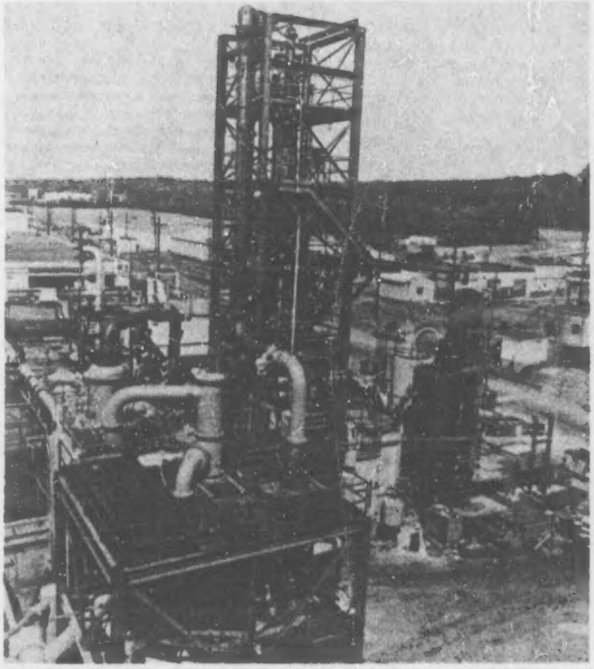


Figure 8. Curtiss-Wright "technology rig" pressurized FBC unit. Initial testing on this unit has been completed, and work has begun on the larger pilot plant. *(Photograph courtesy of Curtiss-Wright Corp.)*

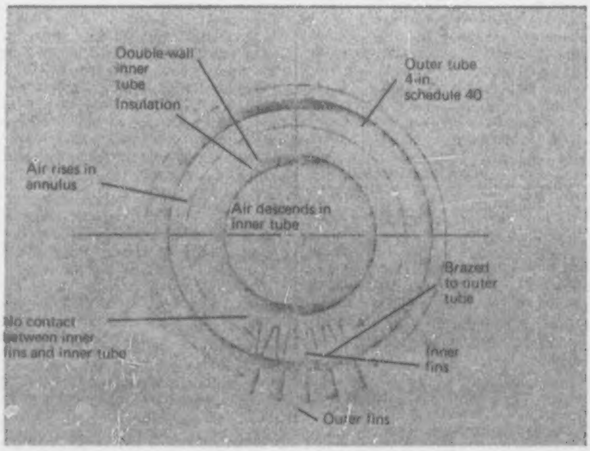


Figure 9. Cross section of concentric tube arrangement used to heat air in the Curtiss-Wright pressurized fluidized bed combustor. The PFBC combustion gases are outside the outer tube.

oil-fired boiler. However, these numbers are constantly changing as fuel prices vary.

IEA project

A subsidiary of the English National Coal Board, NCB (IEA Services) Ltd., is managing a seven-year project to design, build, and test a large PFBC. The United States, the United Kingdom, and the Federal Republic of Germany are sharing equally in the cost of the facility at Grimethorpe in Yorkshire. The International Energy Agency has overall cognizance of this program.

The rating of the combustor is 85 MWt. It is being built as an adjunct to a small power station connected to the National Grid, so all the station facilities are in place, and there is a useful place to use the steam generated.

There are three main objectives for this program:
1. Build the experimental facility to study combustion, heat transfer, gas cleanup, corrosion problems, and energy recovery in a PFBC.
2. Carry out tests over a wide range of operating conditions, and make detailed measurements for these conditions.
3. Provide data that will serve as a base for extrapolation to larger-scale units.

This facility has many of the same features of the units being built in this country. However, since it is primarily an experimental facility, a series of in-bed tube banks will be installed during the course of the experiments. This ability to change tube banks will be matched by the designed-in ability to adjust and rebalance the entire steam/water circuit.

The measurement capabilities of the Grimethorpe facility will be much greater than that of any of the sub-scale or "technology" units in this country. Special arrangements have been made to determine parameters such as air mass flow rate, including employing isokinetic sampling principles. Inlet air is variable over a range of 7 to 31 kg/sec (15 to 68 lb/sec). Large numbers of access ports are located in the walls of the combustor. Gas cleanup uses four primary and four secondary cyclones located around the outside of the combustor shell. The four secondary outputs are manifolded, and the hot gas is routed to the main heat exchanger vessel. Probe holes for dust sampling are located in a number of places in the flue gas cleanup piping. Should there be any significant amount of freeboard burning, which would tend to cause an unacceptable temperature rise in that volume, a special freeboard cooler is provided. If freeboard temperatures continue to rise above 950 C (1800 F) despite use of this cooler, live steam can be injected directly into the freeboard volume.

The combustor vessel is a double-wall device (see Figure 10). Air enters at the top and flows down the annular space. The air is heated by the combustor waterwalls, and it reverses direction at the bottom of the outer vessel and enters the distributor plate. Flue gases leave the top of the combustor and are ducted to the multiple cyclones.

The distributor plate is a bubble cap design, consisting of many small air holes, each about 13 x 2.5 mm (0.5 x 0.1 in.). There are also two 150-mm (6-in.) ash-drain pipes and a 40-mm (1.6 in.) bed material feed pipe that penetrate the distributor plate.

Grimethorpe uses a pressurized steam/water system. Water flows through the walls of the heat exchanger, where it is warmed by the air after it passes through the gas cleanup system. Water then passes through the combustor waterwalls, where it is maintained as single-phase water. Evaporation takes place only in tube banks located in the fluidized bed and in the heat exchanger. All water is turned to steam with a slight amount of superheat. Controls at the exit from the heat exchanger maintain FBC-produced steam at the temperature and pressure of the existing power station steam manifold.

The conditions in and around the in-bed tubes are to be determined during operation, as the combination of fluidizing flow at 2.5 m/sec (8.2 fps) with an overbed pressure of 10 bar (9.9 atmospheres) has not been tested elsewhere.

Startup is the most complicated procedure to be followed in this system. The basic steps are itemized below:
1. The system compressor is operated under local manual control.
2. The required air flow to the combustor is established. A constant fluidization condition must be maintained, regardless of bed pressure and temperature.
3. Bed material is fed under manual control to form a fluidized bed that is initially too shallow to cover the in-bed tubes.
4. Wall-mounted propane burners are ignited. When all are operating properly, propane is admitted to openings under the distributor plate.
5. As soon as proper bed preheat conditions are reached, coal is admitted under local manual control. As bed temperature rises due to burning coal, the propane flow is gradually reduced.
6. Additional coal is fed and propane supply is further reduced until the entire bed is being heated only by combustion of coal.
7. More bed material is fed, and tubes begin to be immersed in the bed. Any cooling effect due to these tubes is countered by increasing the coal feed rate.
8. Bed material and coal are continually added, and air flow is increased, until normal operating conditions are reached. Maximum bed depth during the experimental period will be from 4 to 5 m (13 to 16.5 ft).

The Oak Ridge National Laboratory (ORNL) is heading a team of three companies to develop a program leading to the construction of a test plant at ORNL. This plant will study the use of a coal FBC with a gas turbine for the cogeneration of electricity and industrial process heat. Westinghouse Electric Corp., Concordville, Pa., Babcock & Wilcox Co., New Orleans, La.; and Stone & Webster, Inc., of New York complete the team.

DOE's fossil energy research office's program, Coal Combustor for Cogeneration (CCC), under which this effort will function, is part of a national effort to develop clean-burning coal systems for industry and the electric utilities.

The tasks will include design and construction of a 200-kW test plant; conceptual design for a larger (20-25 MW) plant; preparation of a site-specific demonstration plant design; and preparation for commercialization of the CCC system.

A market assessment conducted by ORNL about a year ago indicates that 61% of the heat used for industrial processes could be supplied by CCC plants. The study also showed that a CCC system can yield heat utilization efficiencies of 60% to 75% and can offer cost savings of as much as 25% over conventional heat sources and purchased electricity.

Babcock & Wilcox Co.

The Babcock & Wilcox Company (B&W) (no connection with Babcock & Wilcox, Ltd.) was recently awarded a contract by TVA to build a 20-MW AFBC pilot plant. In addition to serving as a large-size research tool, this unit will be used to evaluate the operation of a fluidized bed boiler in a utility environment. The unit, to be installed at the TVA Shawnee Steam Station, will have a bed area of 216 ft^2. It will generate steam at typical utility conditions of 2400 psi and 1000 F, and permit testing over a range of fluidizing velocities and bed depths, and of a number of other concepts such as both under-bed and over-bed feeding of the coal and limestone and particulate discharge recycle to improve combustion efficiency and sulfur capture.

B&W is also participating in TVA's design study program for a 200-MW demonstration plant. In this program, TVA is evaluating competing designs from three different contractors, one of whom will be selected to build the AFBC boiler for the demonstration plant. The first phase of this program (conceptual design studies) was completed about 1½ years ago. The program is now in the second phase (final conceptual design) which will be completed in June 1981. The boiler is being designed to generate approximately 1.46 million lb/hr of steam for use in a standard 2400-psi reheat steam cycle.

In addition to the TVA program, B&W has been developing conceptual AFBC boiler designs for a 600-MW unit. This work was begun as a part of the Stone & Webster contract with DOE. This work has continued under the current joint B&W/EPRI AFBC development program where the conceptual design studies are providing direction to the experimental programs.

B&W's AFBC experimental facility includes three units. The smallest, with a 1-ft x 1-ft bed, has been used to evaluate a number of different fuels including various bituminous coals, Texas lignite, oil shale, tar sands, and tar sands coke. The 3-ft x 3-ft bed facility was used in an EPRI program to test parameters affecting limestone utilization.

The third unit, with a 6-ft x 6-ft bed, is a significant element in the joint B&W/EPRI development program. This unit was designed as an intermediate scale test facility to help bridge the gap between the smaller laboratory pilot units and the anticipated utility operated pilot facilities. Since startup in early 1978, this unit has been operated for nearly 3000 hours.

In addition to developing techniques to improve system performance, the operating experience has provided valuable insight into the design requirements for a commercial unit. The highlights of the first 1½ years of operation include the following items:

1. The testing thus far has centered about a fluidizing velocity of 8 fps, a bed temperature of 1550 F, and a flue gas oxygen content of 3%. The combustion efficiency has been in the 92-95% range and the CO emissions have been less than 200 ppm. The high freeboard height

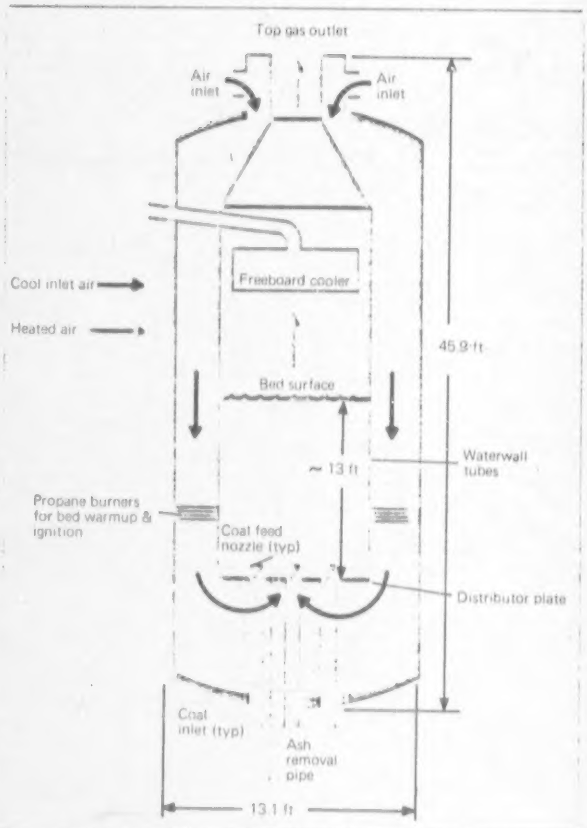


Figure 10. Simplified section through pressure shell of the Grimethorpe pressurized FBC.

(2 sec gas residence time) has been beneficial for both combustion efficiency and sulfur capture.

2. Tests with a single under-bed coal feed point for the 36 ft² bed indicate the potential for increasing the coal feed nozzle spacing beyond the original design of one feed point for 9 ft² of bed.

3. Improved combustion efficiency (approaching 98%) and calcium utilization in the tests with char recycle indicate that recycle may better achieve the desired overall performance than the use of a carbon burn-up bed.

4. Experience with the first two distributor plates (a woven wire mesh plate and a perforated plate) has led the way to the third plate design shown in Figure 11. This bubble cap plate has shown excellent operating characteristics in 1600 hours of testing and has experienced none of the deterioration that plagued the other two designs.

ESCOE research

The Engineering Societies Commission on Energy, Inc. (ESCOE), founded by five major engineering societies, has been doing research on materials of construction for combustion systems, including fluidized bed combustion. The effort on the atmospheric FBC is complete.

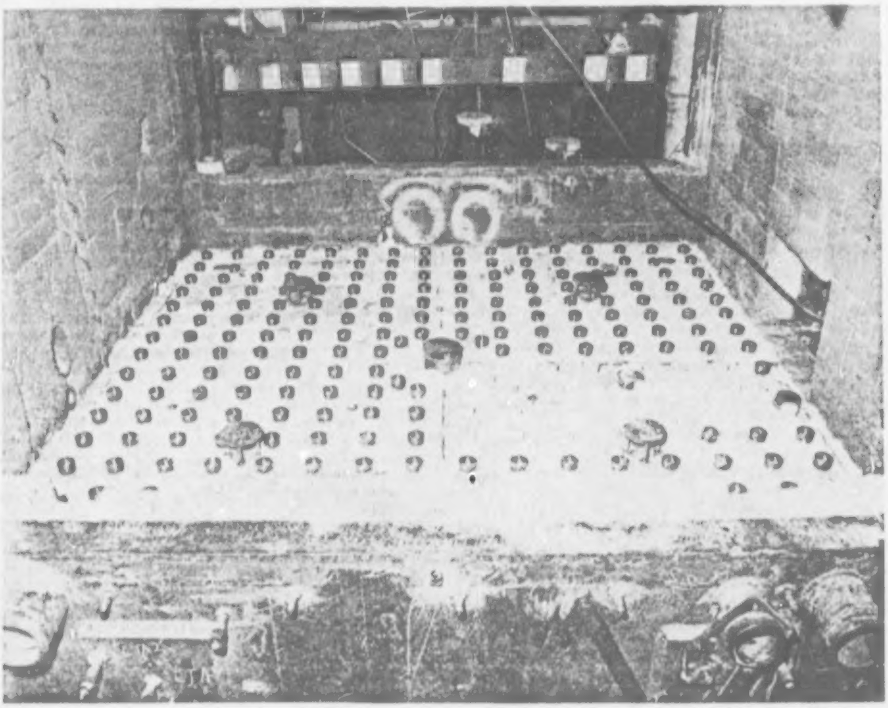
In general, ESCOE indicates, conventional materials of construction can be used in solids handling and peripheral equipment. Above-bed tubes showed little fireside effect when materials were chosen with some care. Tube material usable in the freeboard volume is not necessarily compatible for use within the bed, however.

The fluidized bed, continues ESCOE, is not an equilibrium environment. Materials in the bed are subjected to widely varying conditions. One common problem is pockets of oxygen deficiency, inducing accelerated chemical attack. If these pockets occur in the same bed locations for a variety of bed conditions, these locations could be identified, and no tubes would be placed there.

Bed variables, such as temperature fluctuation, coal feed rate, and fluidizing velocity, may all affect tube corrosion. No complete data exist on corrosion mechanisms in a fluidized bed.

A meliorating factor is an apparent tendency for corrosion rates of in-bed tubing to decrease with operating time. ESCOE indicates this factor should be tested for the long term (up to 10,000 hr) to determine if it is operative.

Figure 11. Babcock & Wilcox FBC air distributor plate with bubble caps atop each air supply pipe. The plate is insulated with refractory material. Warpage has been greatly minimized.



As might be suspected, horizontal tubes are more susceptible to erosion than tubes which are sloped or vertical. The total effect of erosion has been small in most prototype and developmental units, and the only condition under which this may be a problem is if the tubes are located close to the grid plate. In this case, entering gas jets may impinge directly on the tube surface.

Different components of the atmospheric FBC system are subjected to a variety of temperatures:

1. Heat transfer tubing in the bed—metal temperatures up to 1150 F for steam generation at 1050 F; metal temperatures at or near bed temperatures when air is used as the working medium.
2. Support members, perforated plate, parts of the ash drains—at or near the bed temperature.
3. Waterwall tubes in and above the bed—metal temperature near 850 F (may be higher if steam is generated in the waterwall).
4. Above-bed tubes (e.g., superheater, economizer)—maximum metal temperature near 1000 F. Hot combustion gases will not have cooled much before flowing over these tubes.
5. Support members and instrumentation devices in the freeboard—metal temperatures at or slightly below that of the bed.
6. Primary cyclones—gas exiting the flue at the top of the combustor will have had most of its heat removed and should be no higher than 600 F when in contact with the cyclones.

It appears, according to ESCOE, that FBC presents few materials problems except for those components used in heat exchange service directly in the bed. ESCOE recommends a program to determine the best tube material based on long-term testing before commercialized units can be built with confidence and improved economics.

Outlook

What of the future? Is fluidized bed combustion a "new wave" that will make all other methods of steam generation obsolete?

The book *Energy in America's Future* published for Resources for the Future (RFF) by the Johns Hopkins University Press, has caused the energy industry and those who are "energy watchers" to take note because of some of its predictions. The book mentions FBC several times:

"The coupling of fluidized bed technology to cogeneration systems is an attractive prospect for enhanced efficiency in industrial energy use." (p. 165).

"One special advantage of this proposed system (fluidized bed combustion) is that it possibly could be even cheaper to operate than an ordinary coal plant with scrubbers . . . ." (p. 292).

"The new combustion methods, such as fluidized bed combustion, are being designed with the goal of achieving the required cleanup job on coal at modest cost savings over conventional coal plants with scrubbers. Projected cost savings, of course, could be more significant if actual scrubber costs turn out to be larger than expected because of reliability problems with scrubber systems." (p. 300).

" . . . new types of coal technology may meet existing standards for air pollution at less cost (fluidized bed combustion) . . . to get still cleaner air." (p. 301).

" . . . modified methods of coal combustion, such as fluidized-bed combustion . . . are designed so that they can at least meet and often improve upon the emissions levels for sulfur, and usually those for particulates, without using scrubbers." (p. 347).

These comments, virtually the total of reference to utility use of FBC in the text, reflect at best a cautious optimism.

The industrial companies, on the other hand, are most enthusiastic about the prospects of FBC. While they may not see it as a panacea for all the ills of high-sulfur coal burning, they feel that after thousands of hours of testing they are ready to participate in the design, construction, and operation of fluidized systems in a big way.

A recurring theme in the development of FBC is the claim that its use will preclude the need for SO_2 scrubbers. Scrubber technology itself is new and to a degree unproved. Long-term operation and reliability figures on scrubber use are not yet available, especially in retrofit situations. The power generation industry, in its quest to use more coal, is now faced with a difficult choice: to use new scrubber technology, or to use new FBC technology.

It would appear the choice will encompass both technologies. Scrubbers have caught most of the attention, primarily because they can be added on to existing plants without rebuilding the entire boiler section. FBC use is progressing at what apparently is a more deliberate pace. Test and evaluation units are being constructed and operated in a gradual size progression, so unexpected effects caused by large jumps in scale should not occur.

That there is utility interest cannot be overlooked, however. The Rivesville plant is a good start in determining how a unit can work with an operating station. EPRI sponsorship of FBC development permits many utility companies to participate in a limited fashion, reaping common advantages while minimizing individual losses.

But in the main, FBC is a technology that will be accepted slowly by the average utility. Its use may well grow more rapidly in about 10 years. By then, the 200-MW plant design TVA is overseeing should have proved itself; that amount of generating capacity will require a steam rate of approximately 2 million lb/hr. When an FBC installation can provide that much steam on a regular basis, competitive with conventional (1979) technology in both cost and reliability, and without the requirement for an external scrubber, the utilities will no doubt be pursuing fluidized bed combustion most actively.

The sleeper, of course, is the PFBC-turbine cogeneration method. There are two methods here: a hot air/hot gas mixture that drives a gas turbine and a waste-heat steam boiler; and use of hot gas and steam turbines, with both working media being generated in the bed. (This latter is a project of the General Electric Co.) These units, individually small, can be combined to form a station with a capacity in the hundreds of megawatts. Sharing coal and limestone preparation facilities, as well as ash removal and gas cleanup components, will effect considerable efficiencies. A large drawback in this instance is lack of complete knowledge of material behavior in the long term under the conditions of the pressurized fluidized bed. ESCOE, among others, is addressing that matter and will soon publish its findings.

For now, FBC proponents must be satisfied with small industrial units and add-ons to existing fossil-fueled steam power plants. Progress is being made, and the outlook is favorable, but because of today's economic and regulatory climate, FBC may take just a little longer to come to maturity then was first imagined. END

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This article updates the status of AFBC development. Start up of a 20 mw unit by TVA is a major step leading to a large, utility sized facility. Due to technical uncertainties of building a large 200 mw unit, TVA's pilot plant should help to resolve remaining questions regarding the feasibility of future large-scale AFBC.

AFBC design and operating systems to be evaluated in pilot plant

A new TVA/EPRI 20-MW atmospheric fluidized-bed boiler will allow a variety of fuels and combustion parameters to be tested. The design is a scale-up of smaller experimental combustion units

By THOMAS M. MODRAK and J. WILLIAM SMITH, Babcock & Wilcox

Atmospheric fluidized bed combustion (AFBC) boilers have recently become a viable alternative to conventional coal-fueled boilers for industrial power plants. Although in operation in a number of industrial applications, AFBC technology requires further evaluation for eventual scale-up to commercial utility boilers.

A major milestone in the development of this technology for utility boilers was reached in mid-1982 with the completion and dedication of a 20-MW pilot plant by the Tennessee Valley Authority (TVA). The facility, located at TVA's Shawnee Steam Plant on the Ohio River near Paducah, Ky., is the focus of a $28.5-million test program funded jointly by TVA and the Electric Power Research Institute (EPRI). It is designed to evaluate a number of areas of technological uncertainty, most of which relate to size and scale-up considerations, as well as long-term operating characteristics. The basic concept is shown in Figure 1.

A fluidized bed consists of a stream of gas flowing upward through a dense bed of solid particles. As the gas flow increases, bubbles (pockets of air) form and rise through the bed, causing a highly turbulent mixing of bed particles. This action looks similar to boiling fluid and is called a bubbling bed. AFBC takes place in the bubbling bed volume. Principles and examples were discussed in detail in POWER ENGINEERING, November, 1979, "Fluidized Bed Combustion Systems: Progress and Outlook," pages 46-56.

6x6 experimental unit

Babcock & Wilcox (B&W) has built several small test units, with the largest having a combustion area of 6 ft x 6 ft (the "6x6" unit). These were described in the November 1979 article, and the unit is shown in Figure 2.

A major goal of 6x6 unit testing is to increase the efficiency of calcium (Ca) utilization for sulfur (S) capture in the AFBC process. Current scrubber systems for limiting SO_2 emissions offer relatively high efficiency for calcium use in the sulfur capture process. Perfect stoichiometry requires one mole of calcium

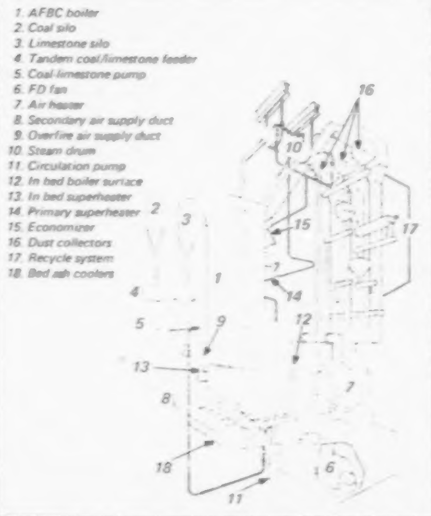
1. AFBC boiler
2. Coal silo
3. Limestone silo
4. Tandem coal/limestone feeder
5. Coal-limestone pump
6. FD fan
7. Air heater
8. Secondary air supply duct
9. Overfire air supply duct
10. Steam drum
11. Circulation pump
12. In bed boiler surface
13. In bed superheater
14. Primary superheater
15. Economizer
16. Dust collectors
17. Recycle system
18. Bed ash coolers
16
10
2
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Figure 1. TVA/EPRI 20-MW AFBC pilot plant. Diagram is simplified to show main components and relative locations.

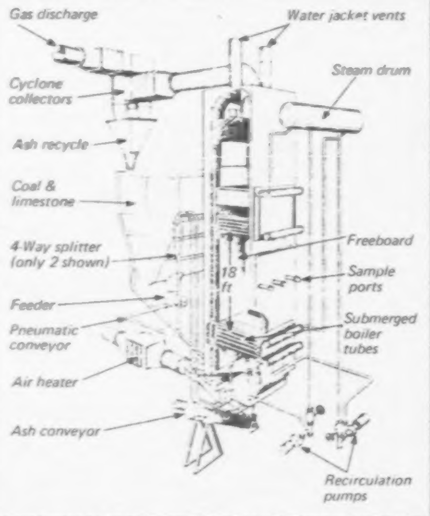
Gas discharge
Water jacket vents
Steam drum
Cyclone collectors
Ash recycle
Coal & limestone
4-Way splitter (only 2 shown)
Freeboard
18 ft
Sample ports
Feeder
Submerged boiler tubes
Pneumatic conveyor
Air heater
Ash conveyor
Recirculation pumps

Figure 2. Babcock & Wilcox/EPRI 6x6 atmospheric fluidized bed combustion test facility.

for each mole of sulfur captured. Scrubber processes require a Ca/S molar ratio slightly in excess of 1.0 to achieve the 90% sulfur capture required for most coals. Without recycle of the elutriated (carried over in the gas stream) bed material, the AFBC process requires up to three times as much calcium (Ca/S = 3 max.) to achieve similar sulfur capture.

This low calcium utilization is a concern for two reasons. First, the cost of limestone detracts from the system's overall economic attractiveness. Second, the high Ca/S ratio affects boiler efficiency. The calcination reaction of the limestone is endothermic (heat-absorbing); the sulfur capture reaction is exothermic (heat-evolving). At 90% sulfur capture, the resulting heats from the two reactions are in balance at a Ca/S mole ratio of approximately 2.5. With high-sulfur coals, boiler efficiency drops approximately 1% for each increase of 1.0 in the mole ratio over 2.5. Therefore, the current goal for the 6x6 development program is to lower the Ca/S ratio required for 90% sulfur capture to 2.0.

During the high-flow-rate particle-carryover recycle tests, where the recycle flow was as great as five times the coal flow, the overall Ca/S ratio required for 90% sulfur capture was still greater than 2.0. A superficial velocity (upward velocity of the hot gases above the bed) in the range of 10 to 15 fps resulted from the test conditions and therefore contributed to the high Ca/S ratio. Work is continuing to improve calcium utilization, with testing at velocities nearer 8 fps and with varying limestone sizing.

20-MW pilot plant

In mid-1979, TVA issued specifications for the design and construction of a 20-MW AFBC pilot plant for testing purposes in support of eventual commercial development. This followed an earlier TVA program which involved the preliminary design of a 200-MW AFBC demonstration unit by several boiler manufacturers. These design studies indicated areas of technical uncertainty, many of which relate to size and scale-up considerations, and to long-term operating characteristics. Rather than incur the risk of immediate scale-up to a 200-MW size unit, TVA decided to construct and operate a large-scale pilot plant for process evaluation.

In October 1979, the turnkey contract for the plant was awarded, and by March 1982 construction necessary to support start-up activities was complete. The first coal firing occurred on May 2, 1982, full boiler load was achieved June 6, 1982, and formal dedication ceremonies for the unit were held August 3, 1982. A cutaway view showing internal details is presented in Figure 3.

Objectives of the AFBC pilot plant are to (1) demonstrate acceptable long-term operation and performance; (2) develop and demonstrate automatic control and

	Design	Variability
Bed area	216 ft^2	
Superficial velocity	8 fps	4, 8, or 12 fps
Bed temperature	1550 F	to 1650 F
Bed depth	4 ft	to 6 ft
Coal feed rate	12,000 lb/hr	to 22,000 lb/hr*
Steam capacity	111,200 lb/hr	to 170,000 lb/hr*
Outlet pressure	2400 psi	2400 psi
Outlet temperature	1000 F	1000 F

* = With low-grade fuel

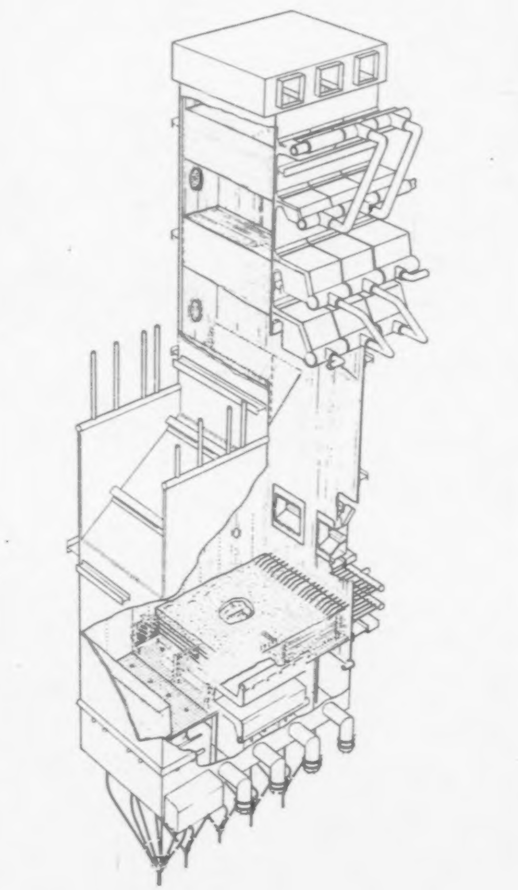

Figure 3. 20-MW TVA/EPRI AFBC pilot plant, with combustion area shown in cutaway view. This unit has 216 ft^2 of active bed area, with a 72-ft^2 start-up area.

load-following capability; (3) develop a data base for multiple fuel and sorbent types to determine design correlations for scale-up to commercial-size boilers; (4) demonstrate the unit's environmental control capability as a basis for assessing the environmental acceptability of AFBC on a commercial scale; (5) develop the capability to specify and design critical auxiliary equipment, especially the coal, limestone, and recycle feed systems; (6) test and evaluate materials for the boiler and its auxiliaries; and (7) train engineers and operators for future units.

To meet these objectives, specifications of the pilot plant required a boiler with design features comparable to those of a commercial-size unit, large enough to minimize uncertainties associated with scale-up, and to operate at utility steam conditions. Bed area was specified at 216 ft², and the unit had to be capable of operation at three discrete, full-load conditions: superficial velocities of 4, 8, and 12 fps, with initial operation at 8 fps. The requirements resulted in a pump-assisted circulation drum-type unit with the design parameters shown in the table.

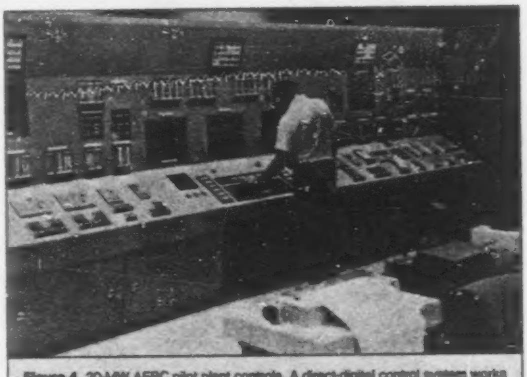

Figure 4. 20-MW AFBC pilot plant controls. A direct-digital control system works with a data acquisiton system to provide on-line data analysis.

The pilot plant site covers 28 acres at the Shawnee steam plant It includes the powerhouse, an office and service building, coal and limestone service areas, handling and preparation facilities, and solid waste storage area. Although located at the Shawnee site, the pilot plant is isolated from the power plant. Also, the pilot plant is designated a 20-MW unit, but it generates no power. The steam produced is condensed to feedwater.

The bed has 216 ft² (12 ft x 18 ft) of active area, divided into four compartments and containing evaporative and secondary superheater tube bundles. An additional 72 ft² (12 ft x 6 ft) of adjacent bed area comprises the startup compartment, where coal combustion begins. No surface is submerged in this compartment.

Two forced-draft fans supply combustion and fluidizing air to the boiler, providing the pressure necessary to force the air through the windbox, distributor plate, and bed. The windbox is divided into nine independently controlled compartments, two for each of the active bed areas, and one for the startup area. This control allows segmental fluidization of the bed for startup and for load turndown testing.

The distributor plate, which also supports the slumped bed when at rest, is of water-cooled membrane-wall construction, with air discharge bubble caps set in the membrane. The tubes forming the distributor plate are integrated into the front and rear walls of the furnace enclosure, so the plate and walls operate at the same temperature. This means that the plate can be welded to the walls, eliminating the expansion and sealing problems which have plagued some AFBC boilers.

The convection tube pass contains primary superheater and economizer tubes. It is divided into three equal gas passes, which allows a constant gas velocity to be maintained through the convection pass while operating at the three different superficial velocities. The design also allows for evaluation of heat transfer characteristics and tube bundle erosion in the pass by varying flue gas velocities and solid loadings.

The 20-MW plant incorporates a number of unique design features which exemplify its function as a research tool. One of these is the use of two separate coal and limestone feed systems—underbed and overbed—in order to evaluate and compare the merits of the two concepts. The underbed feed system supplies the startup and active bed compartments. Three-way splitters in each of the active areas apportion solids to individual feed points that each supply 18 ft² of bed area. Although an underbed system can provide reasonably even distribution of coal and limestone, it is complex and requires long-term, large-scale testing to determine reliability and the maximum bed area which can be supplied by a feed nozzle without affecting system performance.

The overbed feed system includes two spreaders located in the furnace front wall to distribute coal, and a gravity limestone feed system supplying the four overbed feed points located in the furnace sidewall. Overbed feed design is less complex than underbed, but the system requires operational evaluation for AFBC scale-up. There is uncertainty as to the adequacy of fuel distribution and effect on boiler performance and emissions, since a portion of the fuel will burn in suspension before falling to the bed.

The pilot plant is equipped to recycle char collected in cyclone collectors in order to investigate ways to improve combustion efficiency and calcium utilization. Three multiple-tube cyclone collectors, feeders, screw coolers, and storage bins allow for recycling char to the bed. The system is sized to recycle material at up to five times the coal feed rate, and at temperatures varying from 285 F to 800 F. Space is provided to allow the future addition of hot cyclones to recycle 1550 F material.

The testing program for the 20-MW AFBC pilot plant is scheduled to last through mid-1986. The schedule is divided into two parts, the first of which includes four major campaigns as follows: (1) shakedown and baseline configuration testing, 8/82-1/83; (2) feed system selection and evaluation, 2/83-8/83; (3) steady-state turndown and transient evaluation, 9/83-2/84; (4) control system evaluation and improvement, 3/84-6/84.

Testing during part two of the program will be based on earlier campaign results and on the developmental priorities of the utility industry. Potential activities include: testing alternate fuels and sorbents; improving and modifying the control system; streamlining fluid bed combustion auxiliary systems; and studying alternate superficial velocities and high-temperature recycle methods. The control room is shown in Figure 4.

The AFBC boiler promises to become a commercially feasible process to burn coal in an environmentally acceptable manner, at costs competitive with or lower than conventional boiler-plus-scrubber systems. A number of utilities across the country are seriously interested in participating in the construction and operation of a demonstration unit in the 100- to 200-MW size range, the next major step toward AFBC commercialization for utility power generation. END

Three types of advanced Fluidized Bed Combustion (FBC) systems are discussed; circulating beds, multistage units and rotating bed FBC. The first two types have advanced to the commercialization stage in the U.S., while rotating bed processes are under development.

Better fluid-bed units ready to make debuts

Three kinds of designs promise better efficiencies, more flexibility and less corrosion. Commercial-size projects are underway in the U.S. and Europe.

☐ A second generation of coal-fired, fluidized-bed combustors (FBCs) are poised for a shot at the lucrative gas- and oil-boiler market. Several units in the U.S. and Europe (see table) have either made their commercial bows or will do so in coming months. Other designs are either in the feasibility-study or pilot-project phases (the U.S. Dept. of Energy has asked industry to submit proposals for advanced FBC concepts).

The new models, which are mainly of the circulating or multistage variety, are said to provide better combustion efficiency, pollution control and turndown capability[*] than earlier FBCs. In addition, say some manufacturers, the new designs take up less space and are able to solve such known FBC pitfalls as steam-tube erosion and corrosion, and nonuniform heat distribution within the bed. Use of advanced FBCs often allows for simpler coal preparation, note the systems' proponents.

However, despite the high praise, the jury is still out on the new FBCs. Although some large-scale trials have been successful, as yet no commercial advanced-design unit that uses limestone as a desulfurization agent has been proved in operation. And boiler experts have yet to become true believers. For one thing, they note, utilities have had no experience in the operation of the huge cyclones associated with the use of circulating FBCs. As for multistage units, there is considerable potential for fouling.

A STEP UP—The latest advances in FBC technology stem from the need to correct the acknowledged defects of earlier models—steam-tube problems, turndown ratios rarely exceeding 3:1, and average combustion efficiencies of 87-89%.

The novel units improve FBC performance by:

- Putting the steam tubes outside the firing chamber. This separates the heat-transfer and combustor portions of the boiler, reducing erosion and corrosion.
- Establishing separate hot-solids (coal, limestone) and hot-gases exchange systems (Fig. 1). This enables the operator to turn down the hot solids and achieve turndown ratios of 6:1 and higher.
- Recycling solids through the combustion chamber. This allows longer residence times, improving combustion efficiency and pollution control. Since the coal is passed through the system until it burns, there is no need to grind it very finely.

PIONEER WORK—The initial studies of the circulating-bed concept have been done by two European firms: Lurgi Chemie und Hüttentechnik GmbH (Frankfurt, West Germany) and Ahlstrom Co. (Helsinki, Finland). Both have used the advanced models in applications other than coal burning—Lurgi to calcine alumina, and Ahlstrom to burn biomass. However, they

A number of advanced-design fluidized-bed combustors are being developed

Designer	Location	Type	Status	Rating (million Btu/h)	Turndown ratio	Combustion efficiency	SO_2 removal
Lurgi	Lunen, West Germany	Circulating	Startup slated for May 1982	280	3:1	99%	85%
Pyropower	Bakersfield, Calif.	Circulating	Started up in December 1981	50	3:1	98-99%	>90%
Struthers Wells	Uvalde, Tex.	Circulating	Started up last month	50	4:1	N.A.	–
Conoco/Stone & Webster	Lake Charles, La.	Circulating	Awaiting final approval	60	4:1	>94%	–
Wormser Engineering	Dakota City, Neb.	Multistage	Startup slated for late 1982	70	30:1	88-97%	–

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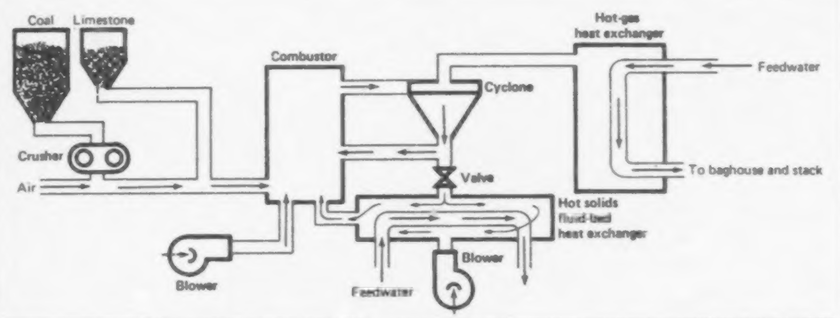
[*] The ability to function at less than full boiler rating. E.g., a boiler with a 3:1 turndown ratio can operate at a minimum capacity of 33% of full rating.


Coal
Limestone
Crusher
Air
Blower
Combustor
Cyclone
Valve
Hot-gas heat exchanger
Feedwater
To baghouse and stack
Hot solids fluid-bed heat exchanger
Blower
Feedwater

Circulating bed allows for higher turndown, less erosion

Fig. 1

have been working with coal/limestone feed for the past several years.

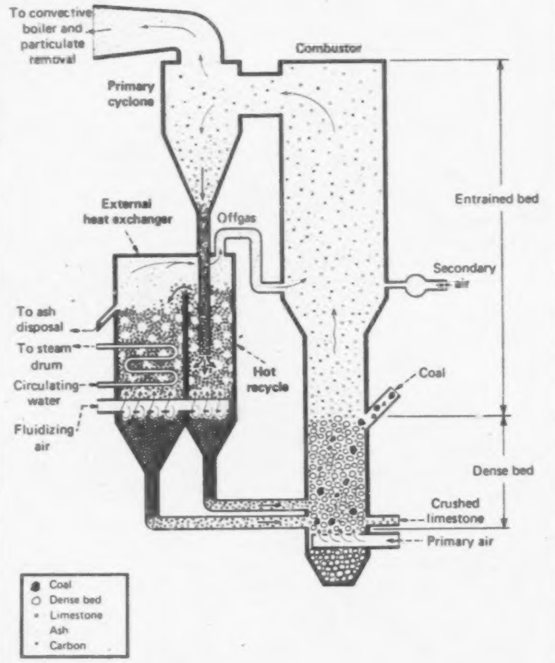
Ahlstrom, for example, has used that feed in a 200,000-lb/h steam cogeneration unit at Kauttua, Finland, that usually is fueled by peat or wood. The tests have been termed successful, but few details are available.

Lurgi has performed more extensive trials at two pilot plants located in Frankfurt. And by mid-summer, the West German concern expects to start up a demonstration/commercial unit for Vereinigte Aluminium Werke at Lünen. The facility, which will be jointly financed by VAW and the West German government, will direct most of its output into making process steam.

Last year Lurgi, along with Combustion Engineering Inc. (Windsor, Conn.) and the Tennessee Valley Authority, designed a 200-MW electric utility circulating-FBC, with an eye toward eventually building a demonstration facility. However, due to reservations about the efficiency of the cyclones, whose size would have reached more than 30 ft in dia., TVA has vetoed further development.

"Cyclones that big just haven't been built," comments a TVA engineer. "We weren't sure we could get enough collection efficiency."

U.S. INSTALLATIONS—Through a San Diego, Calif., firm owned jointly by Ahlstrom and General Atomic Co. (San Diego), one of the first U.S. commercial applications of coal-fired circulating FBCs is going ahead. Pyropower Inc. is installing a 50-million-


To convective boiler and particulate removal
Primary cyclone
Combustor
Entrained bed
External heat exchanger
Offgas
Secondary air
To ash disposal
To steam drum
Circulating water
Fluidizing air
Hot recycle
Coal
Dense bed
Crushed limestone
Primary air
● Coal
○ Dense bed
• Limestone
Ash
· Carbon

Source: Battelle Memorial Institute

Battelle multisolid fluidized-bed combustor

Fig. 2

Btu/h unit, based on the Ahlstrom design, to make steam for an enhanced-oil-recovery project near Bakersfield, Calif., run by Gulf Oil Exploration and Production Co. U.S.A.—a subsidiary of Gulf Oil Corp. (Pittsburgh, Pa.).

At Uvalde, Tex., another circulating-FBC to be used for enhanced oil recovery was started up last month for Conoco Inc. (Stamford, Conn.) by Struthers Wells Corp. (Warren, Pa.). Since last November, Struthers Wells has operated a 5-million-Btu/h model at Winfield, Kan., for the same use.

This design, created by Battelle Memorial Institute (Columbus, Ohio), has what is called a dense bed (see Fig. 2). According to John Fanaritos, a project engineer at Struthers Wells, this feature enhances the combustion of fines, which in conventional FBCs are often blown right out the stack, and in other circulating beds must sometimes be recycled numerous times for complete burning.

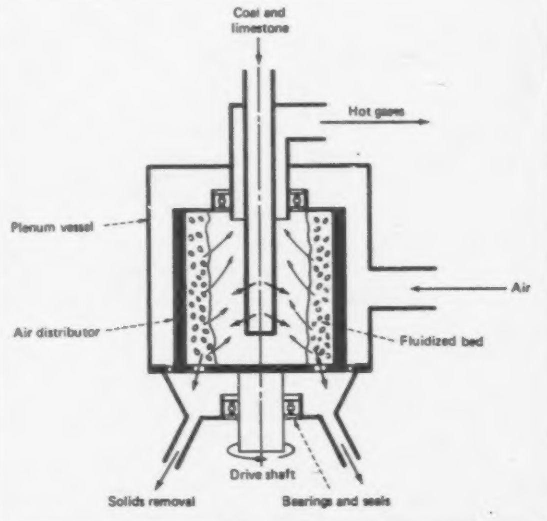
In this system, fines, after going through one cycle, are blocked by the dense bed's barrier of compact limestone and coal, which promotes long residence times. Fanaritos also says that the system burns chunks of up to minus 2-in. mesh, which is larger than what other units burn.

Another circulating-FBC project still has not received the final go-ahead. At presstime, Conoco officials have not decided whether to continue or abandon plans for a 60-billion-Btu/h demonstration model at the firm's Lake Charles, La., chemical complex.

The proposed unit, which has been delayed because of the Conoco merger with Du Pont, would be a scaleup of a pilot system, developed jointly with Stone and Webster Engineering Co. (Houston), and demonstrated last year (*Chementator*, Nov. 30, 1981, p. 17).

On the threshold of marketing a circulating system is a Swedish firm, Studsvik Energiteknik AB (Stockholm). After completing a four-year pilot project, the company is gearing up to offer FBCs of 25 MW and larger to municipalities, for district heating, and to industry. A joint production and marketing agreement is expected to be sealed with boiler maker Generator Industri AB (Göteborg), within the next few months.

ANOTHER APPROACH—Along with circulating beds, the most developed advanced-design FBC is the multistage model, which is closer to conventional FBCs than is the circulating type. While some multistage units have multiple beds, others simply have several air inlets, allowing for staged combustion in one bed.



Rotating-bed design is still far from commercialization **Fig. 3**

Only one system, from Wormser Engineering Inc. (Middleton, Mass.), is currently on the market (see *Chem. Eng.*, Feb. 22, pp. 22-25), but several other groups are exploring the concept. The University of Natal (South Africa) and Howard University (Washington, D.C.) have each come up with design proposals, and the former has built and run an experimental unit.

Some boiler experts maintain that the multistage design has poor combustion efficiencies, caused by brief residence times in the shallow beds (1-1.5 ft deep), and that there is fouling or erosion and corrosion of steam tubes and air distributor plates (of the upper stages). However, Gordon Baty, president of Wormser Engineering, claims that this is not true.

Wormser beds may be shallow, he concedes, but 98-99% carbon burnout is assured, nonetheless, because the unburned coal must pass through an upper bed packed with incandescent limestone. As for deterioration of steam tubes, Baty answers that his company's multistage test units have run for over 1,000 h without encountering such difficulties.

A BED THAT TURNS—Further from commercialization than the circulating or multistage design is the rotating-FBC. Development has gone on at United Technologies Corp. (Hartford, Conn.), Lehigh University (Bethlehem, Pa.) and the U. of Aston, U.K.

In these units, inward drag created by air inflow is balanced by the centrifugal motion of the bed material (see Fig. 3), thus potentially permitting larger airflow and better coal combustion per unit of volume than is possible with a system working against the force of gravity, say the Lehigh researchers.

Rotating beds are a pie-in-the-sky idea right now, points out Floyd W. Crouse, of DOE's Energy Technology Center at Morgantown, W.Va. "It'll be a long time until this gets out of the research phase."

Eric P. Johnson

A fludized bed combustion unit installed at Georgetown University has successfully burned high sulfur coal, while meeting strict emission controls in a city environment. The costs of this FBC process is less than a conventional system with scrubbers.

This installation burns high sulfur coal in an environmentally acceptable manner within a populated area—and works

Fluidized bed combustion at Georgetown University

by Robert L. Gamble
& William J. McCloy

On July 9, 1979, a coal-fired industrial fluidized bed steam generator rated at 100,000 lb/hr was successfully placed into service at the Washington, D.C. campus of Georgetown University. This project, co-funded by Georgetown University and the US Department of Energy (DoE), is intended to demonstrate industrial and institutional application of fluidized bed combustion by burning high-sulfur coal in an environmentally acceptable manner within a populated area. Designed to provide steam for heating and cooling, the unit logged in excess of 6,000 hours through March, 1981. It has consistently operated well within Washington's strict regulations governing emissions of sulfur dioxide (SO_2), nitrogen oxides (NO_x), and particulates.

The Georgetown program has shown that fluidized bed combustion is a promising way to realize the full potential of coal, this country's most abundant energy resource. In a fluidized bed steam generator, a layer of granular limestone is placed on top of a grid designed to prevent the particles from falling through. Air is passed upward through the bed with sufficient force to lift and float the particles. At this point the bed material resembles a boiling liquid. Crushed coal is then introduced into the bed and ignited.

When high-sulfur coal is burned, the limestone absorbs the SO_2 which is produced. The operating temperature of a fluidized bed steam generator is relatively low at approximately 1600°F. As a result, NO_x formation is minimal. Another potential problem encountered in the burning of coal is flyash, a form of particulate. A feature of Foster Wheeler's fluidized bed design is a flyash reinjection system. Flyash is reintroduced into the bed where it is burned again, thereby enhancing fuel efficiency. Nearly all remaining flyash is captured by a "baghouse" dust collector.

At Georgetown, the fluidized bed steam generator and all auxiliary equipment are enclosed in a structure aesthetically compatible with the surrounding university complex. The steam generator has two separate beds which can be fired independently, and operation to date has been with one and two beds in service based on steam load demand. During several operating periods, the fluidized bed unit was the only boiler in service and supplied all the steam required by the more than two million square feet of buildings on the campus. Initially fired with low-sulfur coal, the unit was quickly switched over to the higher-sulfur fuel it will use most of the time.

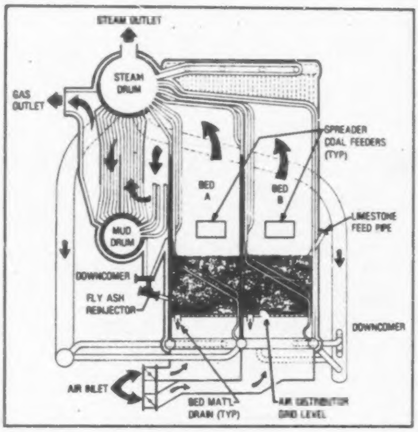


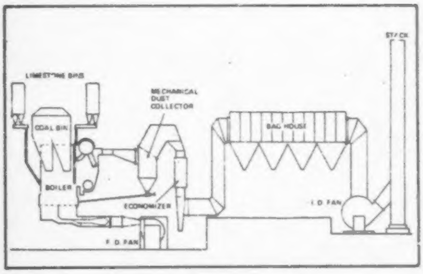
Fig. 1—Cross section of Georgetown installation

Steam generator design

A cross section of the Georgetown installation (Fig. 1) indicates the natural circulation design which can generate 100,000 lb/hr of saturated steam at pressures between 275 and 625 lb/in². Under normal conditions, the unit will be operated at 275 lb/in² in accordance with the university's requirements. The 625 lb/in² operating capability has been provided for future cogeneration application.

Robert L. Gamble is Manager of Development Engineering, and William J. McCloy is Sales Engineer, of Foster Wheeler Energy Corp.

Fig. 2—General arrangement of facility



Each of the independently fired fluidized beds measures 5 ft 6 in. by 19 ft 4 in. Approximately 50% of total unit heat transfer occurs within the 4$^1/_2$ ft deep beds due to excellent heat transfer characteristics. Bed A, adjacent to the boiler bank, is the primary bed into which flyash is reinjected. This bed is always started first and shut down last.

The circuitry shown in part by Fig. 1 includes four external downcomers supplying the lower headers of the unit enclosure and partition wall (separating bed A from bed B) and the sloped tubes within the fluidized bed. Flue gases are cooled by a conventional boiler bank in which the cold-end tubes act as downcomers with the remaining tubes functioning as steam generating risers.

Both beds have separate spreader stoker overbed coal feed systems. Coal, sized to less than 1$^1/_4$ in., is injected into the beds by the overbed spreader feeders. These feeders do not require drying of the coal and have performed well during operation. Sized to less than $^1/_8$ in., limestone is fed by gravity at a single point in each bed. Due to the long residence time of the limestone in the fluidized bed and the rapid burning rate of the coal, less than 5% of the bed consists of fuel.

Since the fluidized bed operates at a temperature of 1600°F, well below the ash softening temperature of the fuel, no slag formation occurs. This eliminates the need for sootblowers as required on other types of coal-fired steam generators. The absence of slagging in the fluidized bed steam generator provides the owner with considerable freedom in fuel procurement practices.

No significant difficulties have been encountered in operating the steam generator, and the unit has performed close to design expectations. Although there are no sootblowers in the system, ash buildup on the tubes in the boiler bank and economizer has not occurred, and cleaning of these areas has not been necessary. Also, due to its dry, friable characteristic, ash carried over from the fluidized beds has not adhered to the heat transfer tubes.

The Georgetown fluidized bed steam generator has already operated briefly at full load, producing 100,000 lb/hr. In addition, full load design capacity of a single bed has been exceeded. With only bed A in service, the load has ranged from 25,000 to 60,000 lb/hr. On the basis of operating experience to date, it is apparent that the bed depth required to make full load is less than expected. This may be due to higher than anticipated heat transfer rates within the fluidized bed or a significant input of heat from the splash zone above the fluidized bed.

Prior to start-up, many concerns about the overbed coal feed system were expressed regarding possible fuel combustion in the freeboard zone above the bed. This problem has not occurred, and freeboard temperatures have been slightly lower than the bed temperature. On a few occasions when coal with a considerable percentage of fine particles has been used, freeboard temperatures have risen to approximately 1800°F. These temperatures were quickly reduced by activation of the water spray system installed in the spreader feeders. With the addition of approximately 5% water by weight to the coal, freeboard temperatures were immediately lowered to below the temperature of the fluidized bed.

Air-gas system

The general arrangement of the Georgetown facility is indicated in Fig. 2. The steam generator is designed for balanced draft operation with air flow from the forced draft (FD) fan regulated by air control dampers for each bed. The FD fan provides sufficient pressure to force combustion air through the air distributors and fluidized bed.

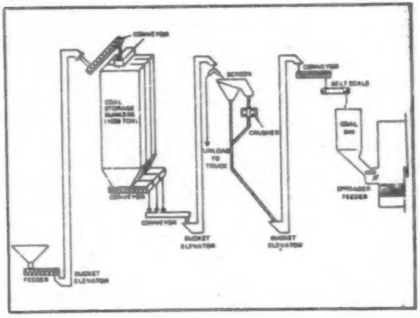
The unit draft is balanced at the surface of the fluidized beds and the induced draft (ID) fan provides the

Fluidized bed projects for West Virginia and Pennsylvania

In addition to the Georgetown unit, a fluidized bed steam generator is located in West Virginia and another is scheduled for a 1981 start-up in Wilkes-Barre, PA. The first, which has successfully generated 300,000 lb/hr of superheated steam is installed at the Rivesville, West Virginia Station of Monongahela Power Company. Described in the July-September 1977 issue of *Heat Engineering*, the Rivesville demonstration facility was designed to power an existing turbine generator at the plant.

The Pennsylvania unit is being sponsored by the City of Wilkes-Barre and US Department of Energy. The design specifies a capacity of 100,000 lb/hr of saturated steam to be used by the city's district heating system. The unit will fire low-quality anthracite culm, a by-product of mining activities which has a heating value of about 5,000 Btu per pound. Approximately 740 million tons are contained in 800 culm banks scattered over the 480 square miles of the Wilkes-Barre anthracite mining region. This is the equivalent of about 1.25 billion barrels of fuel oil.

Fig. 3—Equipment in coal receiving, storage, and feeding system


CONVEYOR
COAL STORAGE BUNKERS
SCREEN
CONVEYOR
BELT SCALE
COAL BIN
CRUSHER
UNLOAD TO TRUCK
SPREADER FEEDER
CONVEYOR
CONVEYOR
BUCKET ELEVATOR
BUCKET ELEVATOR
FEEDER
BUCKET ELEVATOR

necessary suction to draw the flue gases through the boiler bank, mechanical dust collector, economizer, and baghouse dust collector prior to discharge through the stack.

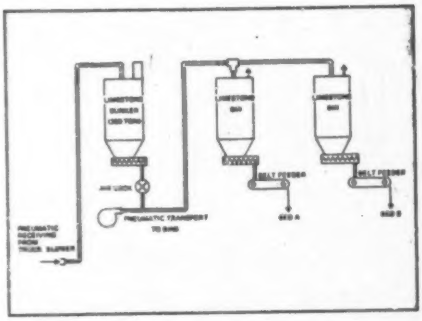
The forced draft and induced draft fans have performed in accordance with requirements and, with the exception of two ID fan bearing failures, have not caused any operating difficulties. The baghouse dust collector has done an excellent job of removing particulates from the flue gas resulting in a clear stack with very low opacity measurements. Stack opacity has been approximately 10%, which is well below the requirement of 20%. Recent analysis of the bags within the baghouse indicates that no acid attack has taken place. Acid attack of the bag materials is one of the primary factors that can limit the life of the bags. The low SO_2 and high calcium content of the ash limits such attack.

Fuel System

To satisfy environmental and aesthetic considerations, coal is delivered to the site in covered trucks and unloaded inside the building which houses the entire facility. The equipment included in the coal receiving, storage, and feeding system is shown in Fig. 3. The coal storage bunkers are designed for mass flow to minimize particle size segregation and hold enough fuel for ten days during full load operation. The coal is passed over a screen and particles larger than 1¹/₂ in. are properly sized with a crusher. Screening the coal before it enters the crusher minimizes generation of fine particles detrimental to the performance of the steam generator. A coal hopper is provided above the spreader feeders on the steam generator and a low level indicator automatically initiates operation of the system to fill the hopper.

The fuel receiving and handling system has functioned essentially without incident. The use of spreader feeders for injecting fuel into the fluidized beds has been especially successful. They have eliminated fuel feeding difficulties experienced on previous fluidized bed units employing pneumatic equipment. Utilization of spreader feeders has also resulted in more than a tenfold reduction in the number of fuel feed points and has greatly increased reliability.

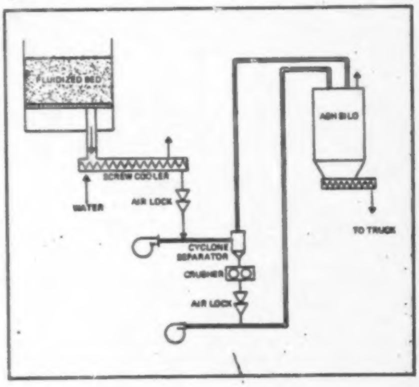
Fig. 4—Limestone delivery


LIMESTONE BUNKER
LIMESTONE BIN
LIMESTONE BIN
BELT FEEDER
BELT FEEDER
AIR LOCK
PNEUMATIC TRANSPORT TO BINS
BED A
BED B
PNEUMATIC RECEIVING FROM TRUCK BLOWER

Limestone system

Limestone presized to less than ¹/₈ in. is delivered to the plant by trucks equipped with blowers which convey the limestone to the bunker shown in Fig. 4. This structure is designed for mass flow like the coal bunkers and holds a ten-day supply during full load operation. Limestone from the bunker is pneumatically transported to bins above each fluidized bed. The discharge from each limestone bin is measured by weighbelt scales before falling by gravity to the beds. Limestone flow is regulated as a ratio of the coal flow, and this ratio is adjusted automatically to control SO_2 emissions at the stack.

Fig. 5—Arrangement of variable speed water cooled screw system


FLUIDIZED BED
ASH SILO
SCREW COOLER
WATER
AIR LOCK
TO TRUCK
CYCLONE SEPARATOR
CRUSHER
AIR LOCK

Removing bed material

Bed material leaves the fluidized bed by gravity through a 6 in. diameter pipe that vertically penetrates the air plenum and air distributor. The flow rate of material from each bed is controlled by a variable speed water-cooled screw device. The arrangement of this system is indicated in Fig. 5.

At the request of DoE, the Georgetown University facility is equipped with a crusher to break up bed material rejects to expose unreacted calcium oxide in the particle cores. This will increase particle reactivity for by-product sale possibilities being evaluated by DoE.

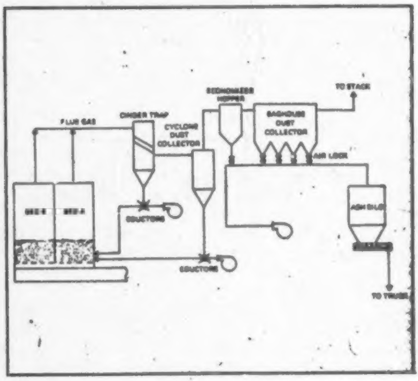
Bed material discharged from the screw coolers is pneumatically conveyed to the crusher and the crushed material is then transported in the same way to a storage silo within the building. Spent material is discharged to trucks within the coal receiving bay for removal from the facility.

As initially installed, the bed material removal system could not be turned down to a rate low enough to facilitate continuous operation. This was resolved by batch operation of the bed material coolers prior to modifying the equipment with a 4-to-1 gear reduction. Since the modification, the removal system has functioned well.

Flyash system

The flyash reinjection system (Fig. 6) has the potential to considerably improve combustion efficiency. Flyash from the fluidized beds is first passed through a cinder trap ahead of the boiler bank to separate coarse particles which are then collected in five hoppers across the width of the boiler. Particles gathered in the cinder trap are reinjected into bed A by eductors—devices that draw the particles from the hoppers and blow them pneumatically into the fluidized bed. Smaller particles are trapped in the cyclone dust collector ahead of the extended surface economizer and are also injected pneumatically into bed A by means of five eductors.

Fig. 6—Flyash reinjection system



Particles not captured in the cyclone unit pass over the economizer to a highly efficient baghouse dust collector. Particles captured in the baghouse are pneumatically conveyed to the ash silo within the building, and this material is periodically discharged to trucks in the coal receiving bay for removal from the facility.

For the most part, the system has performed satisfactorily. However, problems have been encountered with flyash reinjection from the cyclone dust collector. It has been difficult to maintain empty hoppers without excessive transport air in the eductor system. As a result, combustion efficiency in the steam generator has been less than predicted. Steam generator tube erosion has also been experienced on the rear wall of bed A through which the reinjection pipes penetrate. This has been attributed to a bed circulation pattern caused by the high velocity of the flyash reinjection system and the orientation of the reinjection discharge within the bed. Modification of this system is in progress to improve operation.

> ***The safety interlock system will shut the facility down should any critical component fail, or if the bed temperature or furnace pressure does not remain in the proper range.***

Transport and storage of ash from the bed material coolers and baghouse hoppers has been accomplished with few difficulties. The only exception has been the malfunction of drive linkages on the air lock devices below the coolers and hoppers. This has resulted in back flow of transport air into the coolers and hoppers, thereby causing an occasional forced shutdown. These linkage problems are being corrected by the equipment supplier.

Control system

The fluidized bed facility includes a combustion control and safety interlock system that is typical for industrial coal-fired facilities. The system provides automatic control when the load is between 25% and 100%. Operator action is required only at approximately 50% load to light-off or shut down bed B. The safety interlock system will automatically shut the facility down should any critical operating component fail, or in the event that bed temperature or furnace pressure does not remain in the proper range.

Following some initial adjustment difficulties, the fluidized bed facility has successfully operated on automatic control. Further adjustments are being made to maximize the system's rate of response to load changes. The controls are also designed to react to input from an

SO_2 and oxygen gas analysis system. However, reliability problems with the gas analysis equipment have delayed control system adjustments.

Emissions control

Combustion of coal in a fluidized bed of limestone makes it possible to reduce SO_2, NO_x and particulate emissions to levels specified by the Federal government as well as by most state and local authorities. Control of SO_2 is achieved by maintaining the fluidized bed temperature in the range of 1500-1600°F and operating with limestone bed material. The raw limestone ($CaCO_3$) will calcine to lime at the normal bed operating temperature. The lime (CaO) will react with SO_2 released during coal combustion and form gypsum ($CaSO_4$) on the surface of the particles. Affinity for sulfur capture is maintained by continuously feeding limestone and maintaining proper bed inventory with the bed material removal system.

Control of SO_2 is achieved by maintaining the fluidized bed temperature in the range of 1500-1600°F.

NO_x emissions from fluidized bed combustion are low due to the temperature of the fluidized beds. Particulate emissions are kept low by means of a baghouse dust collector designed for 99.75% efficiency. Use of the baghouse has resulted in excellent stack opacity readings and no visible plume from the stack.

Emissions limits for the Washington, D. C. area and those stipulated by the US Environmental Protection Agency (EPA) are summarized by the accompanying table. The Washington regulations also require continuous monitoring of stack opacity in addition to the amounts of SO_2 and NO_x released. As the stack emissions table indicates, performance of the unit at Georgetown has been consistently within these limits. These levels were obtained while firing coal containing 2.3% sulfur.

Table I—Stack emissions

Emission	EPA Limit	Wash., D.C. Limit	Measured (typical)
SO_2 (LB/10^6 Btu)	1.20	0.78	0.46
NO_x (LB/10^6 Btu)	0.70	0.70	0.30
Stack Opacity (%)	20.00	20.00	2.00

Operator training

At the majority of facilities where fluidized bed steam generators will be used in the future, operators will probably lack experience with coal-fired boilers. Since this was the situation at Georgetown, great emphasis was placed on operator training. The training program implemented at the university three months prior to unit start-up played an important part in the overall success of the project.

Classroom discussion, slide presentations, and equipment inspection were employed to thoroughly familiarize operators with the following areas:

- General theory in the design of fluidized bed steam generators, with special attention paid to design details of the Georgetown unit
- Start-up procedures
- Control logic and safety interlock programs
- Maintenance and spare parts program

For more in-depth training, Georgetown University operators assisted the assigned start-up service engineers during final checkout of the boiler and auxiliary equipment. Since initial start-up, service engineering assistance has been continued to give university personnel further guidance in the proper operation and maintenance of a coal-fired system.

Staff requirements include three operators on the day shift and two on the evening and night shifts. Normally, one operator is required in the control room, and one in the plant for equipment maintenance and start-up. During the day shift, a third operator is needed to assist with fuel and limestone receiving as well as with ash disposal.

Classroom discussion, slide presentations, and equipment inspection were employed to thoroughly familiarize operators with various areas.

System performance

Following the resolution of difficulties with the flyash reinjection system, full boiler performance tests will be carried out to determine the system's efficiency. After the first year of operation, the plant's availability has reached the level normally expected for industrial installations.

The Georgetown unit has already shown that fluidized bed combustion of high-sulfur coal can be carried out in a highly populated urban area while meeting strict pollution control regulations. Although some operating difficulties have been experienced, they are typical start-up problems encountered during initial operation of a new facility and are not considered serious. Correction of these difficulties will undoubtedly result in long term reliable operation at good boiler efficiency.

Fluidized bed steam generator systems similar to the Georgetown facility are now being offered commercially. In general, the cost of these systems is less than conventional units requiring flue gas scrubbing for SO_2 removal. □

Gasification combined cycle is a promising means of burning coal cleanly. This EPRI journal article discusses this coal-to-gas conversion process, which is capable of high thermal efficiencies and SO2 removal rates and some of the potential problems in developing full-scale power plants.

A promising technology for converting the fossil energy of coal into electric power is the gasification–combined-cycle (GCC) power plant. It first converts coal to gas, from which sulfur compounds are easily removed, and then burns the clean gas in a combustion turbine at a temperature of 2000°F (1100°C) or higher. A heat-recovery steam generator uses the exhaust gases to power a steam turbine.

In some systems, waste heat from the gasifier can be used to produce additional steam, a step that couples the fuel gas process with the combined-cycle power plant. This emphasizes the essential integration of the fuel gas production and electricity generation processes.

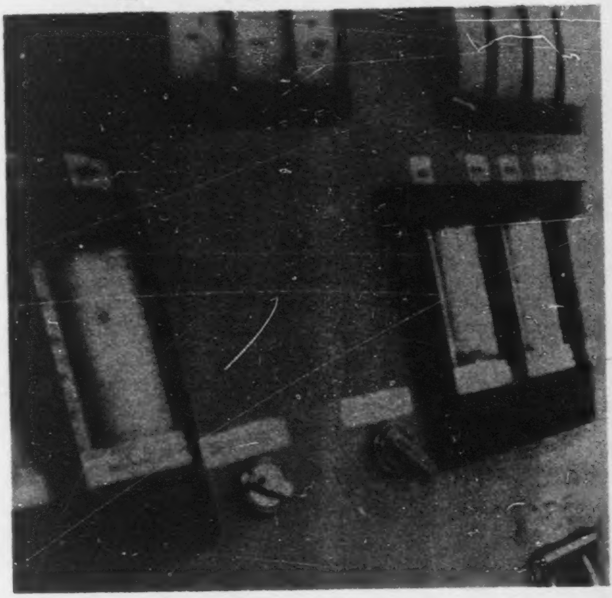
GCC plants will offer many advantages over power plants that burn coal directly to generate steam. They will achieve higher thermal efficiency (lower heat rate), making them economically competitive with other baseload generating options. They will meet emissions standards without serious penalty to plant efficiency and have potential for even lower emissions, if required. Land and water requirements will also be greatly reduced, helping to ease siting restrictions. In addition, GCC plants will respond readily to changes in electric power demand, making them suitable for daily load-following service.

The last-named feature, called maneuverability, was by no means a foregone conclusion. Oil- or gas-fired combined-cycle systems generally respond rapidly and can satisfy daily variations in electrical demand. But the performance of other interacting plant components (such as coal gasification reactors, acid gas absorption columns, and gaseous oxygen plants) had to be determined. In the normal commercial context of gas manufacture, gasification processes operate at a steady pace. It was necessary to show they can maneuver effectively when closely coupled with responsive combined-cycle systems.

To satisfy this important goal, EPRI initiated a program of transient-response studies at pilot plants and detailed simulations of integrated plant operation on

Coupling a Gasifie

Although combined-cycle power plants readily follow variations in load, coal gasifiers were originally designed for steady-state operation. How can these two systems be best linked and coordinated? The answers are coming from pilot gasifier test data and computerized control studies. It appears that fully integrated trains will be able to follow load, besides being cost-competitive for base-load service.



computers. These verified that GCC plants can maneuver to meet load-following requirements. Such analyses are unique in that they established the dynamic performance and controllability of a future process technology well before its commercial development is complete.

The results of these studies are timely as baseline information for the design and operation of GCC plants, most prominently the 100-MW demonstration plant scheduled to come on-line early in 1984 for Southern California Edison Co. at its Cool Water station near Barstow, California. This $300 million project is jointly sponsored by EPRI, Bechtel Corp., General Electric Co., SCE, Texaco, Inc., and others yet to be announced. Its experience will confirm the load-following capabilities of GCC power plants.

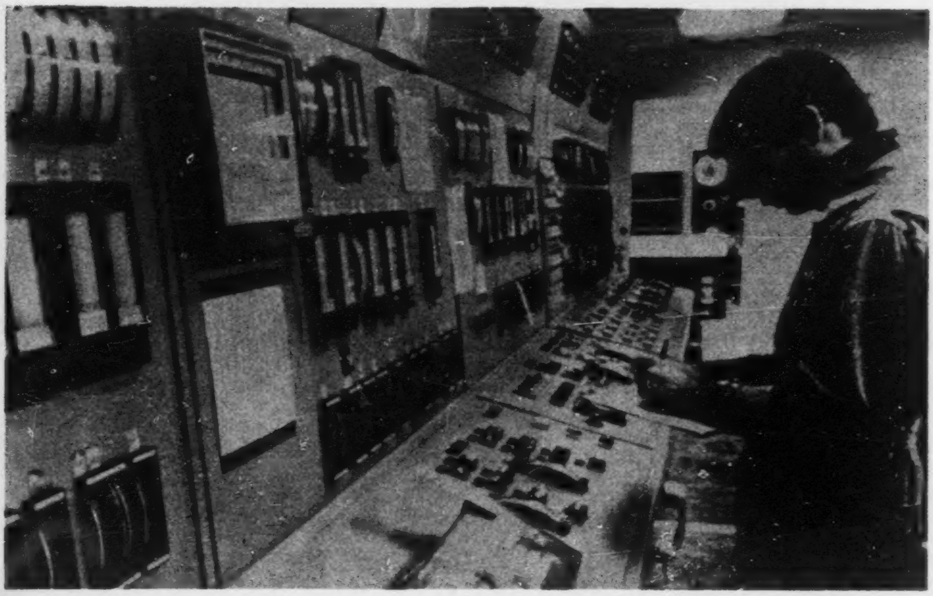
Simulation variables

Large power plants have historically been designed for baseload service. As they age and become expensive to operate, they are relegated to daily cycling with lower annual capacity factors. Therefore, future plants must be flexible—specifically capable of cyclic operation.

It is difficult to set absolute requirements for the response rates of new power plants, but guidelines do exist and were used in the EPRI-sponsored control analyses. Ranges of values are recognized for different operating needs on interconnected power systems. The fastest response rates, 10–20% of unit capacity per minute, are for the very small changes involved in frequency regulation. The slowest rates, 2–5% per minute, are for the 20–40% load swings of daily load following. The guidelines call for a generation unit to be independently stable under all forms of control, controlled to within reasonable response limits, and able to assume a proportionate share of load- or frequency-regulating duty. The criteria imply that even baseload units must be prepared to vary electrical output; units that can load-follow effectively and economically will assume a larger share of the burden in the future.

GCC economic studies for electric power generation generally favor plants

to a Combined-Cycle Plant



that produce medium-Btu fuel gas from oxygen-blown, pressurized, slagging gasifiers of the entrained-flow or moving-bed types. Accordingly, a series of transient-response tests were planned and conducted at three gasification pilot plants: a 15-t/d single-stage entrained-flow plant designed and built by Texaco at Montebello, California; a 165-t/d Texaco plant at Oberhausen, West Germany; and a 350-t/d moving-bed pilot plant owned by British Gas Corp. at Westfield, Scotland. (The last plant is basically a Lurgi, modified by BGC for operation at slagging temperatures.) The pilot test programs demonstrated excellent load-following characteristics.

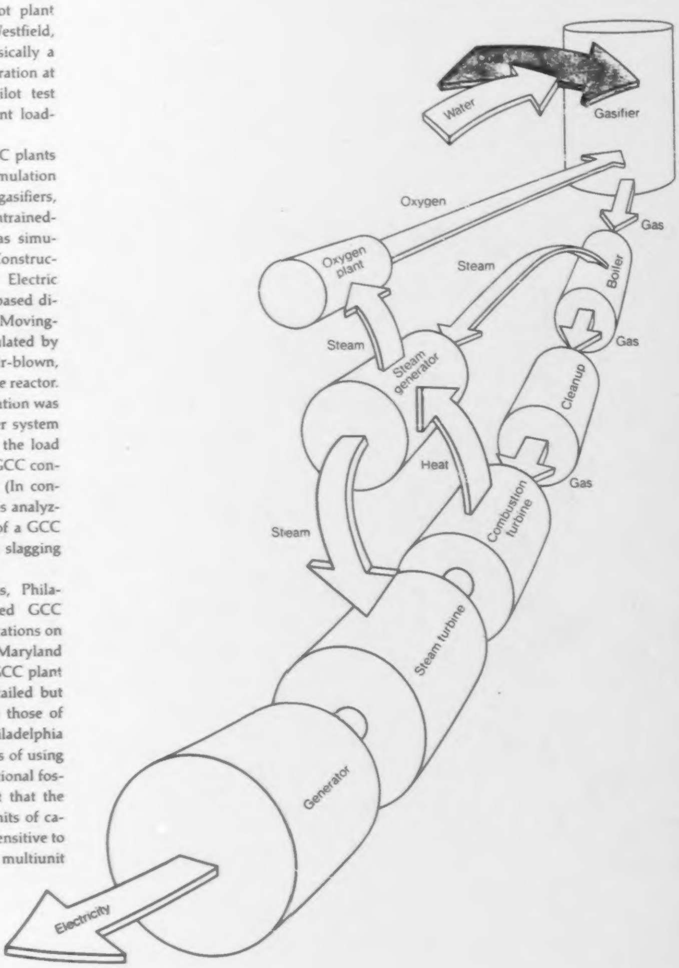
The dynamic response of GCC plants was analyzed in separate simulation studies, also based on different gasifiers, conducted by two contractors. Entrained-flow GCC plant technology was simulated by Fluor Engineers and Constructors, Inc., with Westinghouse Electric Corp. as subcontractor. It was based directly on the Texaco process. Moving-bed GCC technology was simulated by General Electric, based on air-blown, dry-ash operation of a Lurgi-type reactor. In both cases, a full plant simulation was developed for use with a power system network model that portrayed the load patterns needed in evaluating GCC control strategies and capabilities. (In continuing work, General Electric is analyzing the dynamics and control of a GCC plant based on the BGC-Lurgi slagging gasifier in Scotland.)

In more recent simulations, Philadelphia Electric Co. evaluated GCC plants in a model of actual operations on the Pennsylvania–New Jersey–Maryland interconnection. The utility's GCC plant models are somewhat less detailed but correlate reasonably well with those of General Electric and Fluor. Philadelphia Electric is examining the effects of using GCC plants in place of conventional fossil units. Early results suggest that the loading (the order in which units of capacity are brought on-line) is sensitive to the incremental cost curves of multiunit

GCC plant integration schemes are distinguished by the way that steam, air, oxygen, and heat are exchanged between units. Greatly simplified, these are the two plants analyzed in EPRI-sponsored computer simulations of load-following behavior.

The 354-MW entrained-flow gasifier is oxygen-fed and vaporizes water from the coal slurry feed to make its own reaction steam. The 205-MW moving-bed gasifier gets both air and steam from the combined-cycle power unit. The moving-bed gasifier requires considerably more steam to hold its process temperature below the slagging point; extracting this large steam flow from the combined-cycle unit reduces its net power output.

Energy throughput is proportionately greater for an oxygen-fed gasifier because its process stream lacks the large fraction of inert nitrogen found in air. (The product gas is therefore



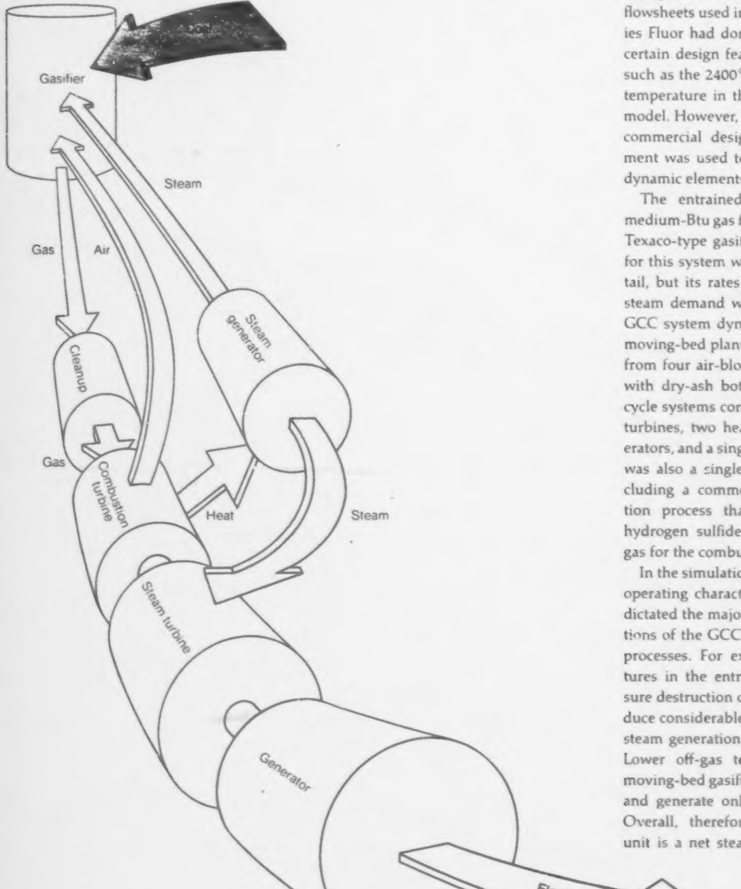
OXYGEN-FED ENTRAINED-FLOW GASIFIER

medium-Btu fuel instead of low-Btu fuel.) This factor results in a smaller volume passing through the gas cleanup unit.

Because the entrained-flow gasifier runs hotter (more than twice the nominal 1000°F, 540°C, of a moving-bed unit), the gas cooling necessary prior to sulfur removal yields significant heat to augment steam generation for the steam turbine. Thus, the ratios of combustion turbine to steam turbine capacity are about 2:1 for the 354-MW entrained-flow GCC plant and about 5:1 for the 205-MW moving-bed GCC plant.

Still under study, a third GCC configuration uses an oxygen-fed moving-bed gasifier modified for operation at slagging temperature. Its medium-Btu product and much lower steam consumption make for better capacity balance between the combustion turbine and steam turbine cycles.

AIR-FED MOVING-BED GASIFIER



plants. These costs are not accurately known because the technology is still under development; however, the model indicates that sequential loading of all units in a given GCC plant (before committing another plant) offers some advantage in response rate.

Process assumptions

For the simulations the GCC plants were configured according to combined-cycle flowsheets used in earlier economic studies Fluor had done for EPRI. Therefore, certain design features were not typical, such as the 2400°F (1300°C) gas turbine temperature in the entrained-flow GCC model. However, in the absence of ready commercial designs, engineering judgment was used to determine the salient dynamic elements.

The entrained-flow plant produced medium-Btu gas from two oxygen-blown Texaco-type gasifiers. The oxygen plant for this system was not simulated in detail, but its rates of oxygen supply and steam demand were varied to influence GCC system dynamics realistically. The moving-bed plant generated low-Btu gas from four air-blown Lurgi-type gasifiers with dry-ash bottoms. Both combined-cycle systems contained two combustion turbines, two heat-recovery steam generators, and a single steam turbine. There was also a single gas cleanup train, including a commercial physical absorption process that selectively removes hydrogen sulfide and yields clean fuel gas for the combustion turbines.

In the simulations, as in the real world, operating characteristics of the gasifiers dictated the major differences in integrations of the GCC plant components and processes. For example, high temperatures in the entrained-flow gasifier ensure destruction of tars and oils and produce considerable heat for high-pressure steam generation in a waste-heat boiler. Lower off-gas temperatures from the moving-bed gasifier result in tars and oils and generate only low-pressure steam. Overall, therefore, the entrained-flow unit is a net steam producer, while the

Load reduction produces distinctive response patterns in the components of entrained-flow and moving-bed GCC plants. These examples illustrate gasifier-lead control and turbine-lead control, respectively. The entrained-flow GCC curves would not change greatly if the control strategies were transposed, but the moving-bed GCC curves would reveal a slower, less stable response.

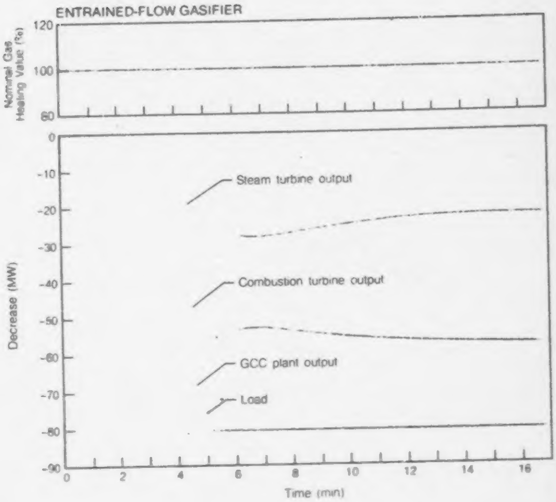
In the 354-MW entrained-flow example, load is ramped down at 4%/min, bringing the plant output down by 80 MW (combustion 57 MW, steam 23 MW). There is a uniform time lag as plant output follows load downward. The plant overshoots slightly at the bottom (about 5.5 min), but the combustion turbine compensates for inertia in the steam turbine response, and within about 10 min the plant is nearly settled at its reduced output level.

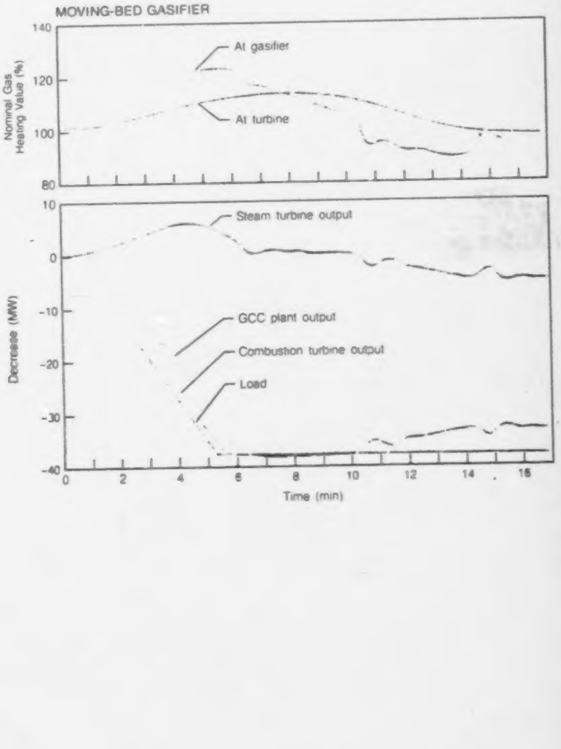
The heating value of the gas (measured at the gasifier exit and, after cleanup, at the turbine inlet) is steady throughout the load change. This is inherent in the entrained-flow process and is the key to the straightforward, unwavering overall response of this GCC plant.

In contrast, when steam and air flows to the moving-bed gasifier are reduced during a turndown of the 205-MW GCC plant, the responses of both turbine cycles are distinctly affected. In this case, the 4% ramp rate cuts plant output by 37 MW (combustion 33 MW, steam 4 MW).

The extra steam momentarily available to the steam turbine causes its output to increase. The plant output proceeds down a slower ramp, and the combustion turbine overshoots its ultimate reduction. Simultaneously, the gasifier production is temporarily enriched (more methane) as coal continues to devolatilize from the hot bed, but with less diluent air and steam.

Thus gas with a higher heating value ultimately (about 7–10 min) reaches the combustion turbine, countering its move toward reduced output. (The brief but very pronounced fluctuations in the curves for this plant occur when some fuel gas must be used to pressurize the coal feed lockhopper.) Stable operation is achieved after about 16 min.





moving-bed unit is a large steam consumer. This directly affects fuel system integration, which, in turn, affects plant response.

The entrained-flow GCC plant offers a higher steam-turbine electrical output because it has additional steam produced in the waste-heat boiler. Conversely, the moving-bed plant requires a large extraction steam flow for the gasifier, resulting in a low steam-turbine electrical output.

Dynamic performance of the two gasifiers is also markedly different. The entrained-flow reactor is capable of large step changes in throughput (coal in, gas out), while holding relatively constant fuel gas temperature, composition, and heating value. This has been substantiated by Texaco's pilot plant tests at Montebello and Oberhausen. By comparison, fuel gas from the air-blown Lurgi-type unit varies widely in heating value during transients. For example, if combustion turbine fuel gas is cut back to follow a lower load, automatic pressure controls immediately reduce the air and steam flows to the gasifier. There is a sudden drop in overall fluid flow in the reaction bed, yet the evolution of volatile hydrocarbons at the top is momentarily unchanged—hence there is a transient increase in heating value. This tends to counter the earlier throttling down; it is thus a positive feedback effect and hampers stable control. Also, in the moving-bed gasifier the gas pressure in the outlet header may drop when the feed coal lockhoppers are being pressurized. This causes variations in the coal devolatilization rate and gas heating value. Comparable effects are not encountered in the entrained-flow gasifier because its coal feed is continuous and the residence time is short.

Subsystems and controls

For purposes of gas cleanup, a packed-column absorber was simulated in these studies, but without a regenerator column or auxiliary equipment. A generalized simulation model of a complete hydrogen sulfide absorption system is being developed, and its dynamics will be examined in detail.

One important finding of the Fluor-Westinghouse simulation was a potentially large mismatch in the rate of change for the entrained-flow gasifier and its oxygen plant. In that model, the partially simulated oxygen plant was permitted only the slow rate of change (10–25% of baseload production per hour) considered typical from earlier experience with high-purity oxygen plant designs. However, later information sheds new light on this point. At EPRI's behest, Air Products & Chemicals, Inc., recently conducted transient-response tests on a 1000-t/d commercial high-purity oxygen plant. Closed-loop computer control was used to attain rates of change fully consistent with the highest rates (10% per minute) expected in an integrated GCC plant.

EPRI's contractors evaluated two basically different GCC control modes, turbine-lead and gasifier-lead. The former is a familiar power plant control strategy: the combustion turbine fuel inlet valve moves in response to changing load on the power system, and the fuel gas flow is regulated accordingly by pressure control of the gasifier feed streams. In the gasifier-lead mode, the gasifier feed valves first respond to changing load, and the combustion turbine valve subsequently responds to pressure control.

Review of both simulations under the two control modes permits some general conclusions. Closed-loop operation was successful in both plants, because the combustion turbines responded rapidly and effectively handled much of the burden.

The fast response rate of the entrained-flow gasifiers, with constant gas heating value, makes this option very suitable for closely coupled operation in a GCC plant. As a result, the entrained-flow plant demonstrated excellent capability for cyclic operation to meet power system needs. Both control modes worked well in closed-loop operation of this overall plant during normal transients and during emergencies after the simulated loss of key components.

As expected, the response rate of the moving-bed gasifiers was slower because of the larger reacting masses of coal. Transient changes in fuel gas heating value also made control of these gasifiers more complex. Operation of the combined-cycle plant, in turn, was accompanied by fuel system transients that inevitably limited the closed-loop response rate during load-changing operations. Nevertheless, in the turbine-lead control mode, this plant was controlled satisfactorily to meet load-following criteria. In the gasifier-lead mode it was less responsive, making this mode less acceptable.

Perhaps the most significant conclusion from these analyses was that both the entrained-flow and the moving-bed GCC plants are capable of closely coupled operation without interposed gas storage buffers or supplementary fuels. On the basis of these simulation studies and the associated experimental work with pilot gasifiers, the load-following maneuverability of GCC plants has now been thoroughly established. Detailed design of the 100-MW Cool Water demonstration plant is proceeding with improved confidence.

Further reading

Economic Studies of Coal Gasification–Combined-Cycle Systems for Electric Power Generation. Final report for RP239, prepared by Fluor Engineers and Constructors, Inc., January 1978. EPRI AF-642.

Entrained-Gasification–Combined-Cycle Control Study. Final report for RP913-1, prepared by Fluor Engineers and Constructors, Inc., and Westinghouse Electric Corp., June–July 1980. EPRI AP-1422 (3 vols.).

D. Ewart et al. "Power Response Requirements for Electric Utility Generating Units." In *Proceedings of 1978 American Power Conference,* Vol. 40, pp. 1139-1150.

Preliminary Design Study for an Integrated Coal Gasification–Combined-Cycle Power Plant. Final report for RP986-4, prepared by Southern California Edison Co., August 1978. EPRI AF-880.

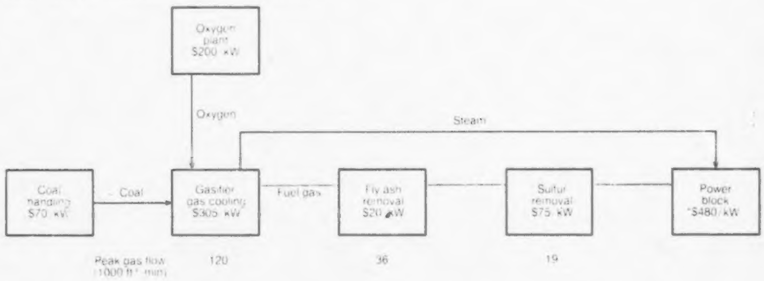
This article, edited by Ralph Whitaker, is based on a paper presented by George Quentin, Advanced Power Systems Division, at the Engineering Foundation Conference on Process Control, Sea Island, Georgia, January 1981.

This recent Electric Power Research Institute (EPRI) report examines the costs of gasification, combined-cycle (GCC) processes and compares these costs with conventional plants with FGD. GCC systems are cost competitive conventional plants, partly due to the lower fuel costs and lower operating and maintenance costs.

COSTS OF COAL GASIFICATION–COMBINED-CYCLE POWER PLANTS

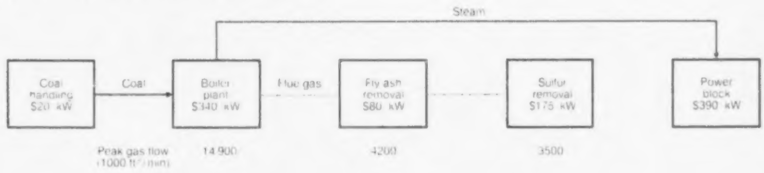
Combustion turbine–steam turbine combined-cycle power plants could have become competitive baseload options if oil and natural gas prices had not increased between 1973 and 1980. Today, such plants can have heat rates of approximately 8000 Btu/kWh (HHV). However, they require ultraclean fuels, such as natural gas and distillate fuel oil, which are now either too expensive or unavailable to the industry as fuels for new baseload plants. An ultraclean fuel can be produced by first gasifying coal by burning it with sufficient oxygen to burn it to combustible carbon monoxide and hydrogen but insufficient to oxidize it completely to carbon dioxide and water. The resulting fuel gas can then be cleaned virtually free of particulate matter, sulfur, and fuel nitrogen compounds. The challenge is to develop an integrated coal gasification–combined-cycle power plant (IGCC) that will be competitive with a pulverized-coal-fired steam plant equipped with required environmental control systems.

Figure 2 compares the estimated costs of various sections of an IGCC power plant with comparable sections of a conventional coal-fired steam plant. The estimates for the IGCC plant were taken from continuing stud-



IGCC Plant With Texaco Gasifier (total cost: $1150/kW)

Oxygen plant $200/kW
Oxygen
Steam
Coal handling $70/kW
Coal
Gasifier gas cooling $305/kW
Fuel gas
Fly ash removal $20/kW
Sulfur removal $75/kW
Power block *$480/kW
Peak gas flow (1000 ft³/min)
120
36
19



Conventional Coal-Fired Steam Plant (total cost: $1005/kW)

Steam
Coal handling $20/kW
Coal
Boiler plant $340/kW
Flue gas
Fly ash removal $80/kW
Sulfur removal $175/kW
Power block $390/kW
Peak gas flow (1000 ft³/min)
14 900
4200
3500

Figure 2 Comparative costs of power plants (1981 $). Total cost for the IGCC plant is $1150/kW when rated at 88°F (31°C); $1005/kW (or the same as the conventional plant) when rated at 60°F (16°C).

ies by Fluor Engineers and Constructors, Inc. (RP239), and represent EPRI's latest estimate of the cost of an IGCC plant using the Texaco coal gasification process and combustion turbines with temperatures limited to those commercially available (~2000°F, 1093°C).

The estimates for the conventional coal-fired steam plant were taken from EPRI report PE-1865, which includes cost estimates by Bechtel Power Corp. Bechtel did not prepare the sectional costs in Figure 2; they were derived by a proration of the Bechtel costs. However, they are believed to be reasonably accurate for this discussion. Bechtel's costs were originally reported in 1978 dollars. They have been escalated here to 1981 dollars using construction cost indexes.

The IGCC system enjoys an initial advantage in the gasifier, fly ash, and sulfur removal sections. This is at least partially understandable by considering the relative gas flow rates through those sections and comparable sections of the conventional plant. This advantage is then offset by the higher costs of other sections of the IGCC plant.

To determine whether an expensive oxygen production plant is feasible for preparing high-purity oxygen for coal gasification when free air can be used for this purpose, the costs of using air, oxygen-enriched air, or oxygen were studied and compared (AF-244 and AP-1624). This work showed that the cost of using oxygen in terms of both capital and product costs was essentially the same as the cost of using the other oxidants. This is fortunate because a nitrogen-free gas can be used to produce products along with electricity that are of interest to the industry, such as methanol and synthetic natural gas. The nitrogen-free gas can also be a superior fuel for integrated fuel cells, such as the molten carbonate fuel cell now under development by EPRI and others.

Most of EPRI's IGCC cost estimates include use of Texaco gasifiers. This is prudent because this gasifier will be used in the 100-MW Cool Water IGCC demonstration plant. In the past year EPRI has begun cost studies of IGCC plants equipped with other advanced gasifiers now being developed for commercial use, specifically, the British Gas Corp.–Lurgi, Shell, and Westinghouse gasifiers. Although conclusions cannot be made at the present time, it is conceivable that at least one of these gasifiers will permit IGCC capital costs to decline.

Another area of potential capital cost improvement is the increase of combustion turbine peak operating temperature (RP1319). If accomplished, this improvement can increase coal-to-busbar efficiency, which may result in an almost proportional reduction in total capital cost per kW of capacity. Reheat combustion turbine development may bring about another possible efficiency increase. A 100-MW demonstration of the reheat combustion turbine is currently under construction in Japan. This plant will not only employ a reheat combustion turbine but is also designed for combustion turbine temperatures as high as 2370°F (1300°C). The Japanese have calculated that this system will have a heat rate of 6800 Btu/kWh (HHV) when fueled with natural gas.

Capital charges represent less than half the busbar electricity costs in coal-fired systems when operated at relatively high annual capacity factors. Thus, although the capital costs of first-generation advanced IGCC plants may be slightly higher than those of conventional coal-fired plants, IGCC plants should enjoy two advantages that will help offset those higher costs. Such advantages could result in lower busbar costs even assuming conventional coal-fired plants will operate at capacity factors as high as IGCC plants. The advantages are lower coal costs and lower total operating and maintenance (O&M) costs. The lower O&M costs result principally from elimination of limestone for sulfur dioxide scrubbing and the cost of scrubber sludge disposal. To date, projections of busbar costs for the two plants operated at the same annual capacity factor indicate that the first-generation IGCC plant can produce power at about a 10% lower cost. (See, for example, AP-1624 and AP-2207.) This cost advantage might increase considerably with improved gasifiers and the advances in combustion turbine technology mentioned above.

Also of importance are advantages that could permit dispatch of IGCC plants at higher annual capacity factors than conventional coal plants.

- Potential higher annual availability
- Lower variable O&M costs because limestone purchase and sulfur dioxide scrubber sludge disposal costs eliminated
- The ability to produce more power at ambient temperatures lower than those used in EPRI cost comparison studies (88°F, 30°C)

Considerable work has been conducted to assess the potential annual availability of IGCC plants (RP1461). The conclusions to date are that these plants can attain equivalent availabilities some 10% higher than conventional coal-fired steam plants primarily because IGCC plant designers incorporate parallel and spare trains. The design philosophy in these designs is to size equipment so that all major items can be shop-fabricated, which reduces the time and expense of field fabrication.

The costs per kilowatt for the IGCC plant shown in Figure 2 are based on an ambient air temperature of 88°F (31°C). Use of this temperature yields performance estimates resulting in pessimistic or worst-case plant costs expressed in dollars per kilowatt. This design basis is used because most design studies are based on plants located in Illinois, a summer peaking area. Recent studies indicate that the net electricity output of these plants will increase appreciably at lower temperatures. For example, a plant with a 500-MW capability at 88°F could produce 550 MW at 60°F (16°C) and 600 MW at 30°F (0°C). The capability is increased because lower-temperature intake air reduces combustion turbine and oxygen plant air compressor horsepower requirements, which results in output increases. Taking the three dispatching advantages into account, the relative year-by-year capacity factors of IGCC and conventional coal plants added to expanding systems are being studied (RP2029-11).

When electricity is generated entirely from steam expansion, heat removed in the condenser is virtually all rejected to the atmosphere by vaporizing water. This continuous water loss means that plants such as conventional coal-fired steam plants consume considerable amounts of water for cooling. Condensers can be air-cooled, which eliminates cooling-water consumption, but at considerable expense. On the other hand, the cost of air cooling an IGCC plant is not considerable because the fraction of steam-cycle power to total power is low (~20–40%, depending on the gasifier). A recent study (AP-2207) indicates that an air-cooled IGCC plant might produce power at lower cost than a water-cooled conventional plant.

Fuel gas produced by coal gasification can be used to generate methanol fuel. Work is proceeding on the development of the unshifted, once-through methanol synthesis process (*EPRI Journal*, June 1982, p. 42). In this synthesis, purified gas from the gasifier is first passed through a methanol synthesis reactor, then on to a combined-cycle power plant. Studies to determine the cost of the produced methanol showed that the methanol would cost considerably less than methanol produced from coal by a nonregulated producer (AP-2212). *Project Managers: B. M. Louks, M. J. Gluckman, and A. E. Lewis*

This paper on cogeneration by a representative of Arkansas Power and Light Company, presents the available benefits of developing this generating technology. The case is made that more efficient energy production will lower costs, decrease fuel consumption and lessen the environmental impacts. Furthermore, since cogeneration systems can operate on a relatively small scale, a variety of fuels could be used.

COGENERATION

A THREAT OR AN OPPORTUNITY FOR THE ELECTRIC UTILITY INDUSTRY

November 18, 1981

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Presented at the

Coal Technology Conference, 1981

COGENERATION

A THREAT OR AN OPPORTUNITY FOR THE ELECTRIC UTILITY INDUSTRY

The merits and potential benefits of cogeneration which are projected to occur because of PURPA has been debated in many forums. A detailed survey is not needed for most of us to reach the conclusion that the majority of electric utilities view PURPA cogeneration as more of a threat than an opportunity.

Rather than dwell on the nature of the prospective threat, the purpose of this paper is to point out what, in the author's judgement, is the greatest opportunity to become available to the electric utility industry since its inception. PURPA was created to assist this nation in acquiring the economic, environmental, and cost saving benefits available through the expanded and extended use of cogeneration and small power production. The Federal Energy Regulatory Commission (FERC) through its "Draft Environmental Impact Statement" for these two sources of power estimates that PUPRA will result in 3400 megawatts of industrial cogeneration by 1995.

Arkansas Power & Light Company (AP&L) is developing a cogeneration program, the use of which makes available benefits far in excess of those PURPA was created to obtain. This program includes:

- Advanced cogeneration technologies with additional fuel saving economics and less adverse impact on the environment.
- New market roles for the members of the three groups who participate in the fuels-energy market.
- Adjustments in energy related legislation and government regulations to allow and aid in obtaining the greater benefits available through the use of this program.

To date, through the use of this program, AP&L has identified in its service area, potential industrial cogeneration of some 1700 megawatts. This AP&L service area potential is more than one-half of the potential forecasted by FERC for the entire nation. This potential will cease to exist if and when the industrial steam users make the switch from oil and natural gas to solid fuel-fired equipment.

For the past thirty to forty years, the roles and market share of the businesses and individuals who operate in the fuels-energy market have been very stable.

There are three basic groups who operate in the fuels energy market:

- Energy and Fuel Suppliers.
- Suppliers of Fuel to Energy Conversion Equipment and Technology.
- Energy Consumers.

In 1973, OPEC forcibly brought to our attention the emergence of a new energy era. Since that time, members of these groups and government have spent large quantities of money and manpower in a diligent effort to maintain the status quo in this market. They have sought to maintain the old energy era status quo by establishing new governmental regulations and legislation and/or developing new technologies as a means of preserving the traditional market roles and market shares. Little attention has been given or resources expended to determine if:

- The forces in this market could forge a more cost effective market structure with new roles and opportunities for those who operate in this market.
- Changes and/or reduction in government regulations and legislation are needed which would encourage and support the market forces to force those who operate in this market into a more cost effective fuels and supply system for this nation.

The author's experience to date indicates that if cogeneration is to be confined to the status quo of the old energy era, the opportunities of using cogeneration to address the problems of the new era are minimal. This experience also indicates that when the concept of new market roles is added to advanced technologies and less restraining governmental regulations, the potential for expanding and extending cogeneration in a manner that is cost effective to this nation is increased 8 to 10 times.

Over the past four years, AP&L has been engaged in assessing the feasibility of extending and expanding the use of cogeneration in its service area. This assessment is an on-going process which has initially addressed the industrial process heat user as the targeted market segment. The majority of these customers have been identified and are for the most part process steam users. Other market segments such as commercial, new industrial installations, residential, pipeline pumping, enhanced oil recovery, etc. are at present scheduled to be addressed at a later time. In addition to the on-going identification of industrial operations who can use cogenerated forms of energy, the marketing and sales concept for addressing this targeted market segment is well along in its development.

The basic cogeneration resources for the targeted market segment are:

- The process heat requirements.
- The electric utility grid.

The process heat requirement takes the place of the electric generation station cooling tower and the grid provides a means of marketing the excess cogenerated energy. The salient feature of the AP&L cogeneration system is a structure which employs new market roles for those who operate in this market and which optimizes the use of the basic cogeneration resources so as to maximize both the fuel saving capabilities and associated economics of cogeneration. The superior fuel saving capability of the AP&L system is set out in Appendix A.

To appreciate the benefits which the use of AP&L's program can provide over the alternatives available to the steam users, an understanding of the following areas is needed:

- The AP&L cogeneration program.
- The alternatives to the AP&L program available to the process heat users and electric utilities.
- The cogeneration potential in AP&L's service area using:
  - The AP&L program;
  - The alternative cogeneration approaches available.

AP&L's alternative to its cogeneration program is the use of the direct combustion of solid fuels in central station power plants. Such plants have heat rates in the 10,000 - 11,000 Btu/kwh range.

The steam users have the alternatives of:

- Continuing to use natural gas and oil in existing process steam only boilers and conventional cogeneration plants. Conventional cogeneration installations (See Figure 1) have heat rates chargeable to power in the 4,000 - 5,000 Btu per kwh range and require some 25 - 50 lbs. of process steam to establish one kilowatt of electric capacity.
- Switching to solid fuel-fired equipment (steam only and/or cogeneration) at existing operations or building new manufacturing operations specifically designed to accommodate the added real estate and other infra-structures that are required for coal fired equipment.

The price of natural gas and oil are expected to continue to rise until at some point in time the price of producing process steam using these fuels in existing equipment (steam only and cogeneration) will exceed the price of steam raised in new coal-fired equipment. To take advantage of this condition and assure that they can compete in their market, the steam users need to have new coal-fired equipment in place shortly after the crossover takes place. Forecasting the time period in which the crossover will occur can be hazardous to the health of corporations and the careers of those involved in this decision-making process. AP&L's present estimate is that natural gas and #6 oil will cost industrials some $10 - $11 per million Btu's in the 1985 - 1987 timeframe, and that the price differential in Arkansas, between these petroleum based fuels and coal at that time will make steam produced in new coal-fired equipment competitive with steam produced with natural gas and oil in existing facilities.

Wood is a solid fuel which will receive some consideration for the smaller installations in AP&L's service territory, but the primary solid fuel for both steam only and conventional cogeneration will be coal. Coal is a fuel which is dirty, difficult to produce, transport, and burn. As compared with oil and natural gas, its use in direct combustion increases the adverse impact on our environment. The use of coal involves engaging in undesirable activities and produces undesirable conditions as follows:

- New investments in equipment which do not increase the industrial plant's output.
- Significant adverse impact on the environment due to the mining, transportation, and burning of coal.

- Lack of qualified personnel who now are not in the labor market. Training must be instituted for individuals who have manual skills and are proficient in areas of technical knowledge related to the use of coal.

- Compliance with environmental regulations which are presently uncertain and as the nation becomes more crowded, promises are to continue to be uncertain and restrictive. These regulations and their uncertainty result in potential additional expenses of unknown amounts.

- Compliance with uncertain environmental reporting and monitoring regulations while the coal-fired equipment is in operation.

- Compliance with uncertain environmental regulations, reporting, and monitoring requirements and their associated costs at the ash and scrubber sludge disposal sites for an indefinite period after the coal-fired equipment is taken out of service.

The work to date on the AP&L program indicates that the required technology is, or shortly will be, commercially available. It also shows that in the 1985 - 1987 time frame the system can produce process steam at prices and conditions which compete with steam produced by the other alternatives available to the steam user. The AP&L program is described generally as follows (See Figure 2):

- Solid fuels are gasified in central gasification installations (See Figure 3) to produce synthetic medium Btu gas (MBG). The synthetic gas is cost-effectively pipelineable some 150 - 200 miles.

- The central gasification installation(s) is located to minimize the delivered cost of feedstocks and sized to take advantage of the associated economies of scale.

- The MBG is pipelined to industrial operations which have process heat requirements where it is used to fire combined cycle cogeneration plants (C^3).

- The C^3 plants are designed and operated to supply the industrial thermal requirement as the primary manufactured energy product. The electricity produced is the secondary product and is taken to the AP&L grid.

- In appropriate configurations, the C^3 plants are applied to repower (replace the natural gas/oil fired boilers in existing conventional cogeneration plants, Figures 4 and 5) or to replace natural gas/oil fired boiler installations which only produce process steam (See Figures 6 and 7).

- The C^3 plants have heat rates chargeable to power in the 5,000 - 6,000 Btu/kwh range, and require only 3 to 5 pounds of process steam to cogenerate a kw of electric capacity or one kwh (See Figures 4 and 6).

- The C^3 plants are configured to use multiple combustion turbines/waste heat recovery boiler units, supplementary firing of the waste heat recovery boiler(s), and on-site standby fuel storage of,

#2 oil initially and methanol at a later date, so as to provide process steam availability and reliability equal to or better than that provided by direct fired boiler installation with redundancy (See Figures 8, 9, & 10).

- C^3 plants are configured so the steam consumer's need for steam can be decoupled from the electric utilities dispatcher's need to control the electric output (See Figure 11).
- Provides for the prompt installation of the C^3 plants when customer's schedule requires steam prior to the completion of the gasification installations. These units would initially be fired on natural gas and #2 oil to provide a bridging arrangement until the medium Btu gas was available at the steam user's manufacturing operation.
- AP&L proposes to own, maintain, and otherwise be responsible for all the facilities required to supply the process heat user's manufactured energy requirements. When some form of joint ownership of all or part of the facilities is productive and mutually advantageous, AP&L desires to explore the use of such arrangements.
- The cogeneration system (gasifier-C^3 plants) would be owned by an unregulated subsidiary(s) and thermal energy sold by such subsidiary(s). The electricity would continue to be supplied under utility regulations.
- Changes in legislation to place electric utilities on equal footing with other cogeneration approaches has been proposed by AP&L (See Appendix B).

A program for the identification of industrial operations with process heat requirements has been underway for several years and is an on-going process. The majority of these operations have been identified and are set out in Table I. Figure 12 is a map of Arkansas and the cross-sectioned part indicates the area within the state where these installations are located. The process steam requirements of these potential steam customers are set out in magnitude order in the first column in Table II.

These requirements cover only the steam produced by imported fuels. They do not include steam raised by fuels produced on site (for example, steam from the black liquor boiler in a paper mill). These installations have an aggregate requirement of some 6 million pounds of steam per hour (pph). They are, for the most part, continuous flow process plants. Steam usage and steam flow profile information is difficult to obtain, but a well-educated estimate of the composite annual load factor is 80 percent.

The fuel required to raise this steam ranges from 1200 - 1500 Btu's per pound of steam, depending on pressures, amount of condensate return, etc. This usage and the forecast fuel costs at \$10 - \$11 per million Btu's translates into a prospective composite steam cost of \$500 million to \$600 million per year in the 1985 - 1987 timeframe.

Six of these installations presently raise their steam requirements with conventional cogeneration installations and cogenerate some 230 megawatts of electric capacity as set out in the third column in Table II. They account for approximately 60 percent of the 6 million pounds per hour total.

The fifth column in Table 2 lists the technical potential for additional cogenerated electric capacity through the application of conventional cogeneration to the process steam requirements set out in the first column. This potential of some 120 megawatts is roughly half of the 232 MW of existing cogeneration capacity. The fourth column sets out the potential for additional cogenerated electric capacity when the AP&L program is applied to supplying the process steam requirements tabulated in the first column. The use of the AP&L program produces a potential of some 1700 megawatts which is approximately 14 times greater than the additional potential of 120 MW from the use of conventional technology. This proposed system not only supplies fuel (derived from coal) for 6,000,000 pph of process steam and 1700 MW of new cogeneration capacity but also the existing 230 MW or a total of 1,930 MW of cogeneration capacity. Although 100 percent market penetration is not likely, if the full potential were to be realized, a gasification-C^3 plant system with the following capacities, inputs, and outputs would be required:

- o Process steam production capacity of 6 million pph and some 42 million thousand pounds of steam produced each year.
- o 1700 megawatts of electric generation capacity distributed across South and Central Arkansas and producing some 11.9 million kilowatt hours per year.
- o Gasification and MBG pipeline system with 460 billion Btu's per day capacity and producing some 142,715 billion Btu's of MBG per year.
- o Gasification feed stock of some 28,120 tons of mid-western coal per day and some 8.7 million tons per year.

This system is capital intensive. The gasifier and pipeline system is estimated to account for 75% and the C^3 plant(s) the remaining 25% of the investment. Assuming the unlikely attainable 100% market penetration and an optimistic four (4) year construction period ending January, 1986, an estimated $5.2 billion would need to be expended during the construction period.

The author has concluded that the electric utility industry is the only institution which can provide the leadership necessary to establish AP&L type cogeneration programs. Cogeneration facilities have in the past been installed by energy intensive continuous flow process industries whose process steam profile reasonably tracks their electric energy usage profile. The steam conditions at the turbine generator throttle were selected so that the electric capacity of their conventional cogeneration installation reasonably matched the load imposed by the electric equipment in their manufacturing installation. PURPA encourages existing cogenerators to optimize their conventional system to maximize electric production by providing a means of marketing such electric energy. This same encouragement is given to steam users who are not now cogenerating. Optimizing the production of electricity in conventional cogeneration plants usually involves the incremental cost and incremental technology change associated with raising the temperature and pressure ratings of the boiler and steam turbine. While this change may call for retraining personnel in changing operating procedures it is not a major departure from past practices for companies with existing conventional cogeneration facilities.

Industrials have little reason to be interested in pursuing the AP&L type cogeneration system on their own when production of some 8 to 10 times as much electric energy they need on site requires almost as much resources in money, management, and technical manpower as is needed to enter a new business. Rather than the incremental addition of investment and technology change associated with upgrading conventional technology, implementation by industry of AP&L type programs involves large investments as well as the need to address unfamiliar and advanced technologies.

For example, the existing cogenerators (See Table 2) raise some 3,840,000 pph of steam from imported fuels. Assuming the use of C^3 plants which exactly match this process steam requirement, some 1,011 megawatts of electric capacity would result. Gasification capacity of 1,011,000 x 12,020 x 24) or 292 billion Btu's per day (See Figure 4) would be required and would result in investments made during a four (4) year construction period ending January, 1986 of some $3.4 billion.

Through discussion with these companies, the author has concluded that while they are willing to investigate some sort of joint venture in the AP&L program, rather than replace existing oil and gas boilers with coal fired equipment, they have shown a preference for using available capital for production equipment rather than new coal fired equipment where they can be assured of purchasing a reliable cost-effective supply of steam.

These conditions leave the possessor of the other cogeneration resource (the electric utility grid) as the only institution with a vested interest in developing and implementing AP&L type cogeneration programs. In the author's judgement, this national resource for energy conservation, the industrial process heat requirement, will be frittered away through the lack of development or the use of conventional cogeneration technology unless the electric utility industry takes the leadership in developing and implementing AP&L type programs before these industries make the switch to coal fired equipment.

Implementation of AP&L type programs by electric utilities provide a means of not only expanding their opportunities for acquiring benefits but also providing the opportunity for other businesses and elements of government to obtain benefits. Some of the opportunities and potential benefits which give the utilities a vested interest in developing and implementing such plans are as follows:

- An opportunity to expand their product line from a single product, electricity, to a multiple product line which includes several forms of manufactured energy.

- With an expanding product line comes the opportunity to address that segment of the energy market in their service area which is composed of the stationary energy consuming devices of various industrial and commercial institutions.

- Sales in new markets bring new revenue streams which the electric utilities badly need. In 1980, AP&L's gross revenue was $750,497,000. The $500 - $600 million potential revenues from the sales of steam in the 1985 - 1987 time frame becomes an expressive figure when compared to the 1980 gross revenue figure.

- Retention of existing electric customers with process heat loads who are candidates for the application of cogeneration. Such candidates usually have high process steam and electric load requirements. Retention of such customers also results in the retention of associated revenues and avoids deterioration of system load factor and the resulting adverse impact on the remaining electric customers.

- Diminishing need for central station power plant.
- New generating capacity distributed over the service area.

The steam users benefit in that they avoid problems and expenses associated with the direct combustion of coal. Capital slated for direct combustion coal fired equipment is free to finance new production equipment. Development and implementation of new training programs associated with coal fired equipment is avoided. The site of the manufactured facility is not encumbered with coal piles, ash, and scrubber sludge disposal areas, etc. In general, such programs offer the potential of supplying the industrial operation with its manufactured energy requirements and frees its management and monetary resources to pursue its primary market.

The use of AP&L type programs results in less adverse impact on the environment than the direct combustion at industrial operations and central station power plants. The gasification/C^3 power plant approach turns the disposal expenses and problems of direct combustion into by product marketing opportunities. The entrained gasifier produces inert ash which, if it cannot be sold for aggregate or fill material, can be disposed of without the expense of perpetual care required for ash from direct combustion systems. The sulfur is removed as elemental sulfur for which there is a market. The nitrogen has a potential market or can be vented to the atmosphere at no cost. The by product cogenerated steam can be sold or used to drive the oxygen plant. The MBG can be made as clean or cleaner than natural gas. The potential adverse impact areas are the production of NO_x at the C^3 plant and sludge resulting from cleaning the water in which the molten ash to gasifier is quenched. These are not serious problems and technologies are available for their control and safe disposal at reasonable cost. A mental comparison of the environmental problems of direct combustion systems to raise steam at 35 industrial operations and a 1700 megawatt central station power plant in comparison to the gasification/C^3 system is sufficient to indicate the magnitude of the difference.

The construction of AP&L type systems gives an opportunity to pay ourselves for the production of manufactured energy rather than paying OPEC for oil from which to produce such energy. The author has not yet investigated the impact on our economy of implementing these programs and leaves to the reader to mentally assess the impact on new jobs, new taxes (federal, state, and local), less unemployment, the need for welfare programs, and the general ability of our nation to compete in the world market.

The author does not wish to give the impression that the development and implementation of such programs is an easy task. Quite to the contrary, the task is expected to be very difficult. If it were an easy task, such systems would be in common use. The time remaining before industrial steam users make a commitment to direct combustion of coal fired equipment is difficult to assess. The lingering natural gas bubble and stabilized oil prices can extend it while an assassination or brush war in the Middle East can hasten its closing. When this opportunity window in time closes, the opportunity and associated benefits are lost for the 20 or 30 year life of the coal fired equipment. Hopefully we will find a way to address this opportunity before the window closes.

TABLE I
PROSPECTIVE CUSTOMERS FOR AP&L-PRODUCED COGENERATED ENERGY
(As of 1/26/81)

	CUSTOMER	LOCATION
1	Alcoa	Benton, Arkansas
2	Anthony Timberlands, Inc.	Bearden, Arkansas
3	Arkansas Chemicals Inc.	El Dorado, Arkansas
4	Bear Brand Roofing, Inc.	Bearden, Arkansas
5	Berry Petroleum Company	Stephens, Arkansas
6	Burlington Industries	Monticello, Arkansas
7	Calion Lumber Company, Inc.	Calion, Arkansas
8	Cooper Industrial Products	El Dorado, Arkansas
9	Country Pride Food Limited (Inc.)	El Dorado, Arkansas
10	Cross Oil & Refinery Co. of Arkansas	Smackover, Arkansas
11	Deltic Farm & Timber Company, Inc.	Waldo, Arkansas
12	The Dow Chemical Company (Inc.)	Magnolia, Arkansas
13	Elk Corporation of Arkansas	Stephens, Arkansas
14	Ethyl Corporation	Magnolia, Arkansas
15	Firestone Coated Fabrics Company	Magnolia, Arkansas
16	Georgia-Pacific Corporation, Crossett Div.	Crossett, Arkansas
17	Georgia-Pacific Corporation, Crossett Div.	El Dorado, Arkansas
18	Georgia-Pacific Corporation Hardwood Mill	Fordyce, Arkansas
19	Georgia-Pacific Corporation	Stamps, Arkansas
20	Great Lakes Chemical Corporation	El Dorado, Arkansas
21	Great Lakes Chemical Corporation	Marysville, Arkansas
22	International Paper Company (Inc.)	Camden, Arkansas
23	International Paper Company (Inc.)	Pine Bluff, Arkansas
24	MacMillan Ring-Free Oil Company, Inc.	Norphlet, Arkansas
25	Manville Forest Products Corporation	Huttig, Arkansas
26	Masonite Corporation	Warren, Arkansas
27	Monsanto Company (Inc.)	El Dorado, Arkansas
28	Potlatch Corporation	Warren, Arkansas
29	Potlatch Corporation	McGehee, Arkansas
30	Potlatch Corporation	Warren, Arkansas
31	Reynolds Metals Company	Benton; Arkansas
32	Superwood Corporation	North Little Rock, Ark.
33	Velsicol Chemical Corporation	El Dorado, Arkansas
34	Weyerhaeuser Company, Paper Mill	Pine Bluff, Arkansas
35	Willamette Industries, Inc.	Emerson, Arkansas

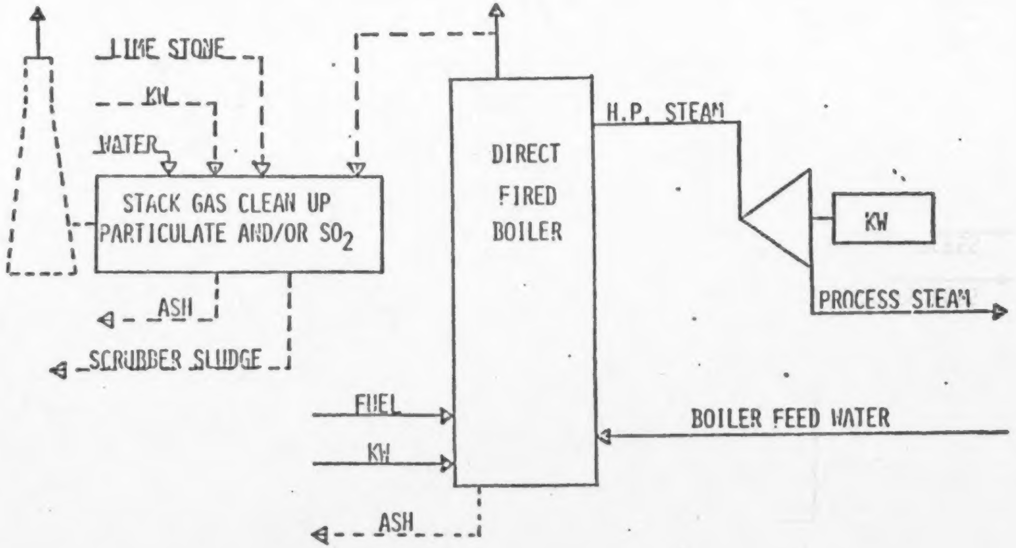
1/27/81

INDUSTRIAL CUSTOMER REQUIREMENTS

PROCESS STEAM	PURCHASES FROM AP&L	PRESENT COGENERATION (CUSTOMER OWNED)	POTENTIAL NEW COGENERATION (ABOVE PRESENT COGENERATION)	
			AP&L OWNED	CUSTOMER OWNED
PPH	KW	KW	KW	KW
1,100,000	8,000	80,000	290,000	0
950,000	12,060	20,000	315,000	12,000
800,000	6,760	30,000	300,000	0
375,000	8,808	20,000	100,000	0
364,000	22,599	10,000	95,800	16,200
250,000	40,067	72,000	62,500	25,000
250,000	12,774	------	62,500	8,800
240,400	1,941	------	60,000	8,000
200,000	1,274	------	50,000	7,000
200,000	6,486	------	50,000	7,000
120,000	5,344	------	30,000	3,000
100,000	1,371	------	25,000	4,000
100,000	5,252	------	25,000	4,000
100,000	6,184	------	25,000	4,000
89,000	2,936	------	22,250	2,225
80,000	5,262	------	20,000	2,000
75,000	1,917	------	18,750	1,875
70,000	2,631	------	17,500	2,000
70,000	2,984	------	17,500	1,750
70,000	1,588	------	17,500	1,750
55,000	4,113	------	13,750	1,375
50,000	2,900	------	12,500	1,250
45,000	3,644	------	11,250	1,125
45,000	844	------	11,250	1,000
35,000	5,895	------	8,750	921
35,000	1,638	------	8,750	700
35,000	3,256	------	8,750	820
28,000	2,089	------	7,000	560
25,000	2,483	------	6,250	500
22,850	461	------	5,700	457
20,000	4,240	------	5,000	455
12,000	1,961	------	3,000	290
10,000	432	------	2,500	200
9,500	562	------	2,375	190
9,500	995	------	2,370	200
6,040,250	191,751	232,000	1,713,495	120,641

FIGURE 1

CONVENTIONAL DIRECT COMBUSTION COGENERATION


LIME STONE
KW
WATER
STACK GAS CLEAN UP PARTICULATE AND/OR SO2
ASH
SCRUBBER SLUDGE
DIRECT FIRED BOILER
H.P. STEAM
KW
PROCESS STEAM
BOILER FEED WATER
FUEL
KW
ASH

1. LBS. OF STEAM TO PRODUCE ONE KWH (DEPENDS ON DELTA STEAM CONDITIONS OF PROCESS STEAM AND H.P. STEAM AT TURBINE THROTTLE). = 25 - 50 LBS. PER HR.
2. HEAT CHARGEABLE TO ELECT POWER = 4000 - 4000 BTU/KWH

FIGURE 2

GASIFICATION - COMBINED CYCLE COGENERATION

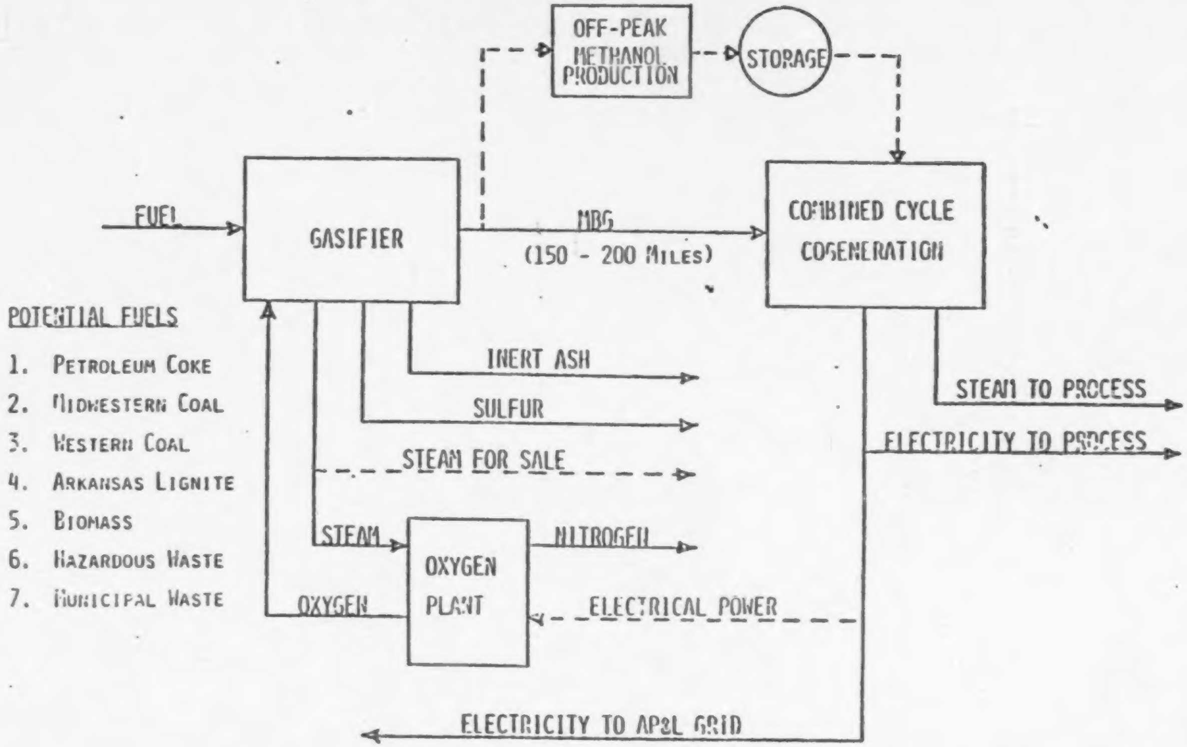

OFF-PEAK METHANOL PRODUCTION
STORAGE
FUEL
GASIFIER
MBG
(150 - 200 Miles)
COMBINED CYCLE COGENERATION
INERT ASH
SULFUR
STEAM FOR SALE
STEAM
NITROGEN
OXYGEN PLANT
OXYGEN
ELECTRICAL POWER
STEAM TO PROCESS
ELECTRICITY TO PROCESS
ELECTRICITY TO AP&L GRID
POTENTIAL FUELS
1. Petroleum Coke
2. Midwestern Coal
3. Western Coal
4. Arkansas Lignite
5. Biomass
6. Hazardous Waste
7. Municipal Waste

FIGURE 3
COAL GASIFICATION

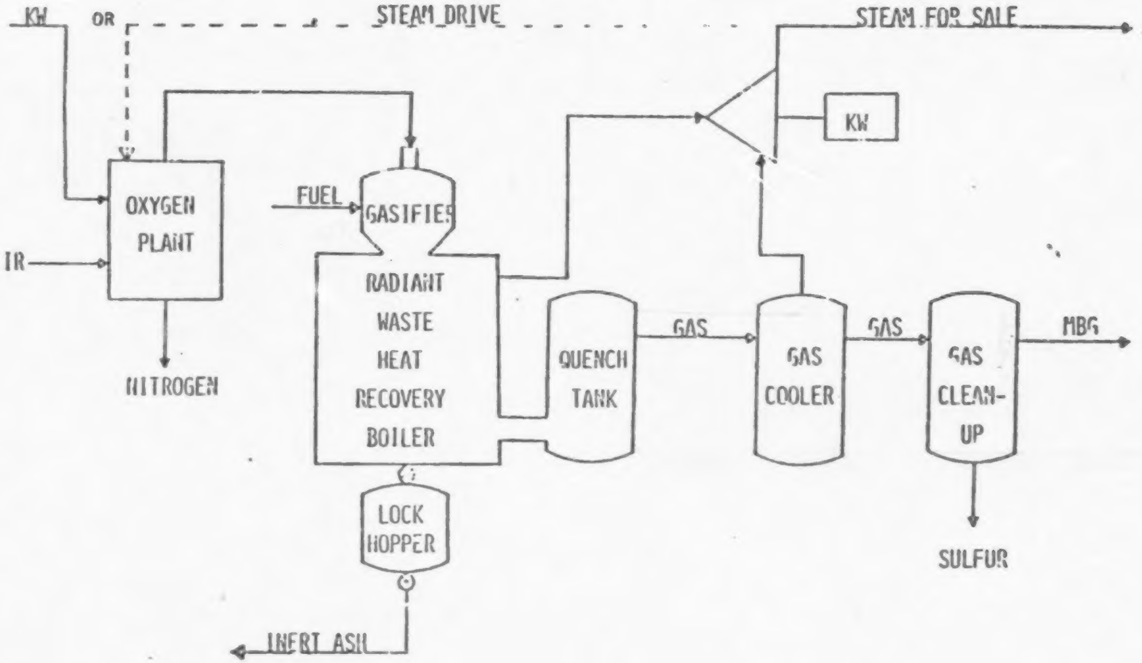
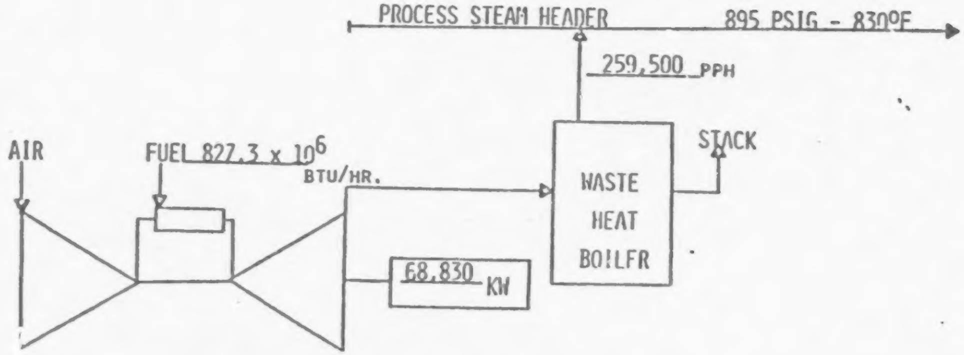

KW
OR
STEAM DRIVE
STEAM FOR SALE
IR
OXYGEN PLANT
NITROGEN
FUEL
GASIFIER
RADIANT WASTE HEAT RECOVERY BOILER
LOCK HOPPER
INERT ASH
QUENCH TANK
GAS
GAS COOLER
GAS
GAS CLEAN-UP
MBG
SULFUR
KW

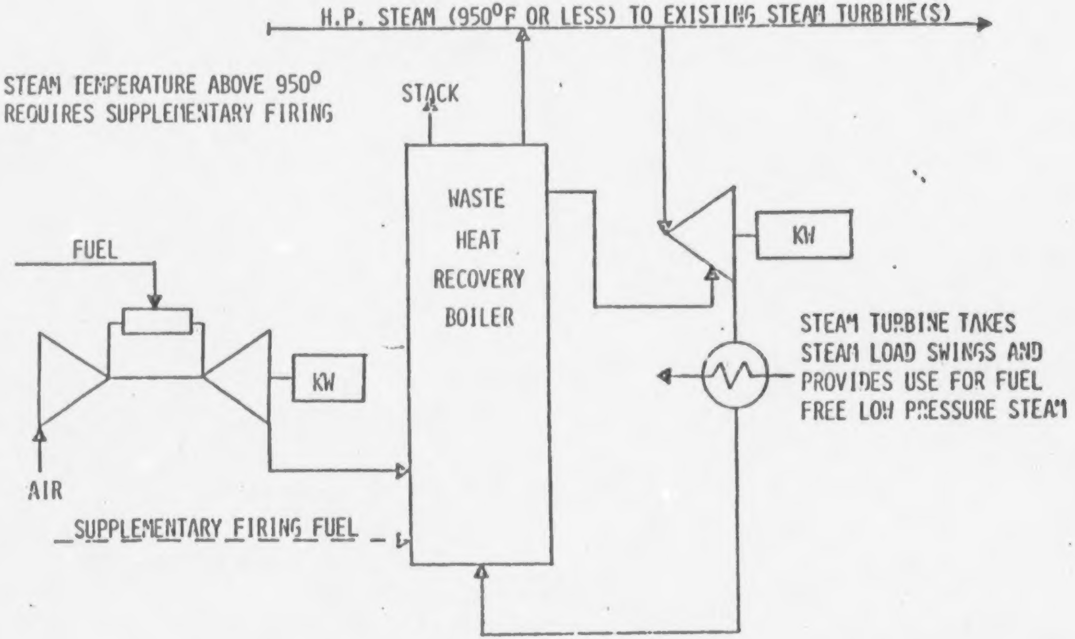
FIGURE 4
COMBINED CYCLE COGENERATION
GENERAL ELECTRIC MS7001 (E) NATURAL GAS


PROCESS STEAM HEADER
895 PSIG - 830°F
259,500 PPH
AIR
FUEL 827.3 x 10^6 BTU/HR.
STACK
WASTE HEAT BOILER
68,830 KW

1. LBS OF PROCESS STEAM TO PRODUCE ONE KWH = 3.8 LBS/KWH
2. FUEL TO PRODUCE ONE KWH AND 3.8 LBS OF PROCESS STEAM = 12,020 BTU/KWH
3. ELECTRICAL OUTPUT (GROSS) = 68,830 KW

(NOTE: DATA FROM G.E. BULLETIN GER-2229N)

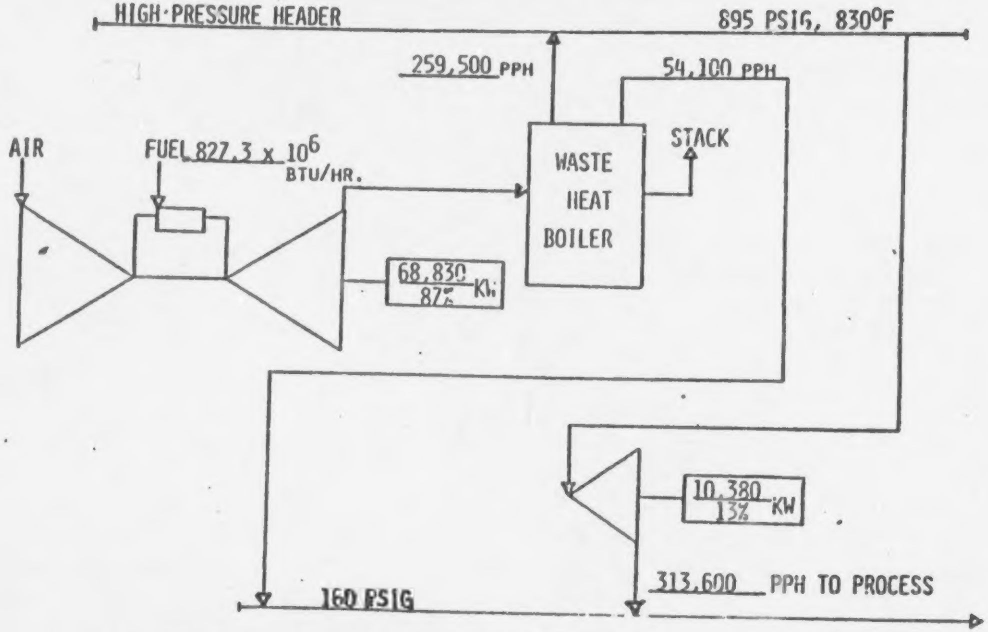
FIGURE 5
COMBINED CYCLE COGENERATION
FOR
REPOWERING EXISTING CONVENTIONAL COGENERATION PLANTS


H.P. STEAM (950°F OR LESS) TO EXISTING STEAM TURBINE(S)
STACK
WASTE HEAT RECOVERY BOILER
FUEL
KW
KW
AIR
SUPPLEMENTARY FIRING FUEL

NOTE: STEAM TEMPERATURE ABOVE 950°
REQUIRES SUPPLEMENTARY FIRING

STEAM TURBINE TAKES
STEAM LOAD SWINGS AND
PROVIDES USE FOR FUEL
FREE LOW PRESSURE STEAM

FIGURE 6
COMBINED CYCLE COGENERATION
GENERAL ELECTRIC MS7001 (E) NATURAL GAS


HIGH·PRESSURE HEADER
895 PSIG, 830°F
259,500 PPH
54,100 PPH
AIR
FUEL 827.3 x 10⁶ BTU/HR.
WASTE HEAT BOILER
STACK
68,830 KW 87%
10,380 KW 13%
160 PSIG
313,600 PPH TO PROCESS

1. LBS OF PROCESS STEAM TO PRODUCE ONE KW = 4.0 LBS/KW
2. FUEL TO PRODUCE ONE KW AND 4.0 LBS OF PROCESS STEAM = 10,400 BTU/KWH
3. ELECTRICAL OUTPUT (GROSS) = 79,210 KW

(NOTE: DATA FROM GE BULLETIN GER 2229H).

FIGURE 7

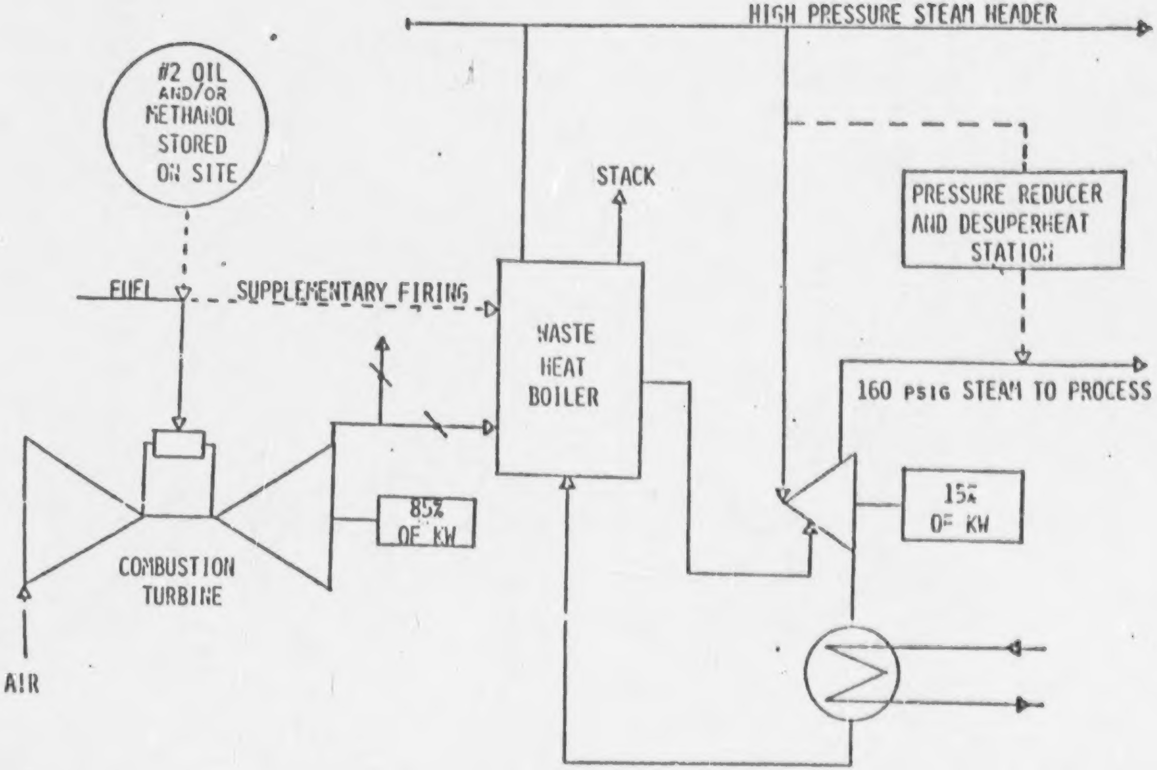

HIGH PRESSURE STEAM HEADER
#2 OIL AND/OR METHANOL STORED ON SITE
FUEL
SUPPLEMENTARY FIRING
STACK
PRESSURE REDUCER AND DESUPERHEAT STATION
WASTE HEAT BOILER
160 PSIG STEAM TO PROCESS
85% OF KW
15% OF KW
COMBUSTION TURBINE
AIR

FIGURE 8

FOUR GENERAL ELECTRIC'S MS7001

NO FAILURES

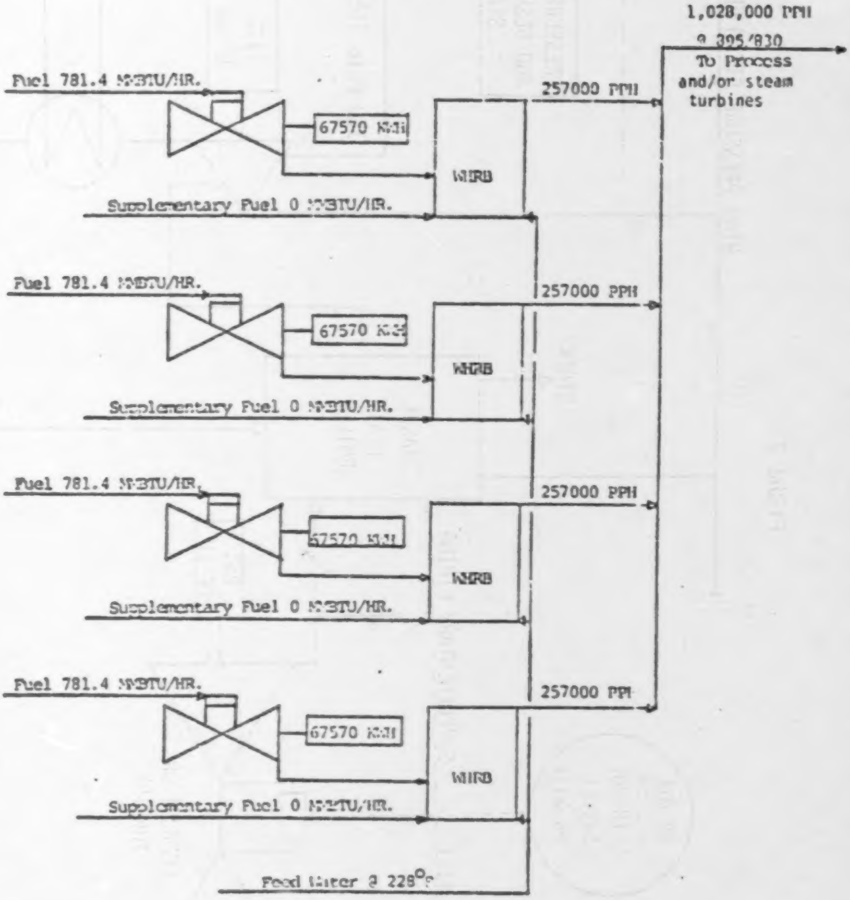

1,028,000 PPH
@ 395/830
To Process
and/or steam
turbines
Fuel 781.4 MMBTU/HR.
67570 KW
WHRB
Supplementary Fuel 0 MMBTU/HR.
257000 PPH
Fuel 781.4 MMBTU/HR.
67570 KW
WHRB
Supplementary Fuel 0 MMBTU/HR.
257000 PPH
Fuel 781.4 MMBTU/HR.
67570 KW
WHRB
Supplementary Fuel 0 MMBTU/HR.
257000 PPH
Fuel 781.4 MMBTU/HR.
67570 KW
WHRB
Supplementary Fuel 0 MMBTU/HR.
257000 PPH
Feed Water @ 228°F

FIGURE 9

FOUR GENERAL ELECTRIC'S MS7001

FIRST CONTINGENCY FAILURE

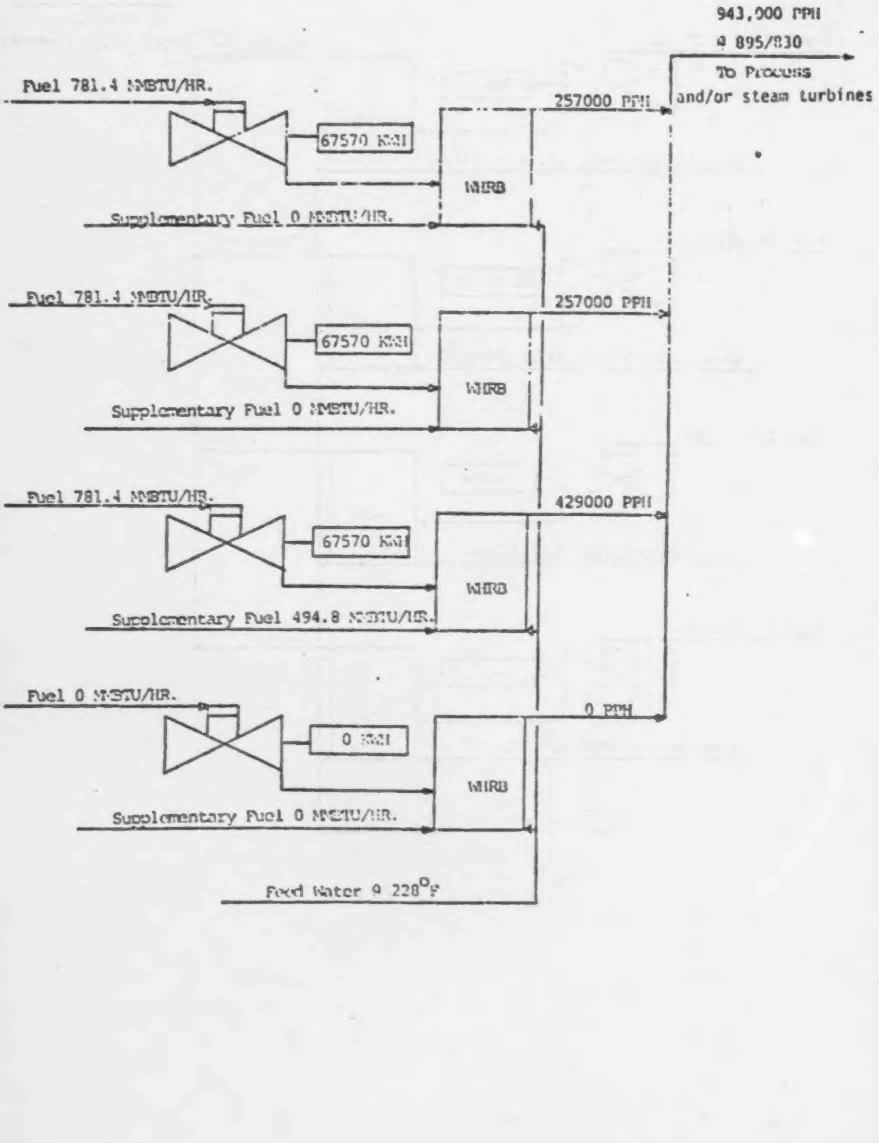

943,000 PPH
@ 895/830
To Process
and/or steam turbines
Fuel 781.4 MMBTU/HR.
67570 KW
WHRB
Supplementary Fuel 0 MMBTU/HR.
257000 PPH
Fuel 781.4 MMBTU/HR.
67570 KW
WHRB
Supplementary Fuel 0 MMBTU/HR.
257000 PPH
Fuel 781.4 MMBTU/HR.
67570 KW
WHRB
Supplementary Fuel 494.8 MMBTU/HR.
429000 PPH
Fuel 0 MMBTU/HR.
0 KW
WHRB
Supplementary Fuel 0 MMBTU/HR.
0 PPH
Feed Water @ 228°F

FIGURE 10

FOUR GENERAL ELECTRIC'S MS7001

SECOND CONTINGENCY FAILURE

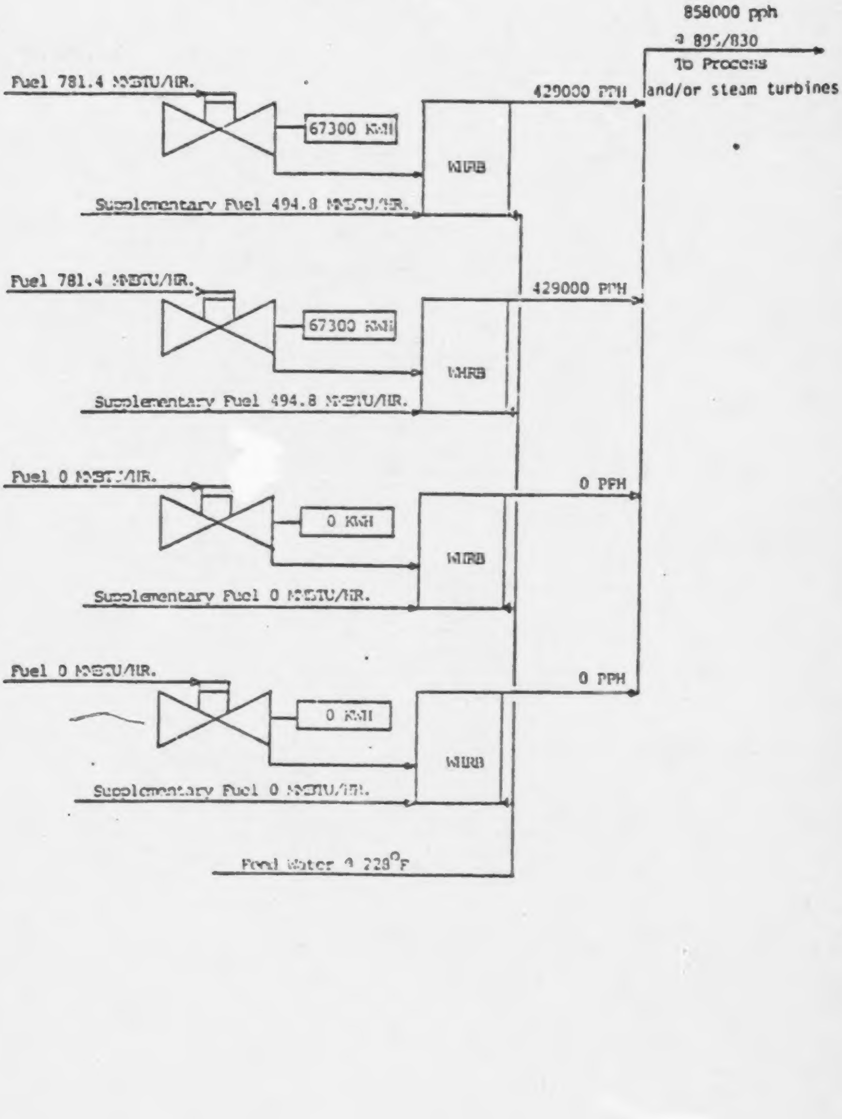

858000 pph
@ 895/830
To Process
and/or steam turbines
Fuel 781.4 MMBTU/HR.
67300 KWH
WHRB
Supplementary Fuel 494.8 MMBTU/HR.
429000 PPH
Fuel 781.4 MMBTU/HR.
67300 KWH
WHRB
Supplementary Fuel 494.8 MMBTU/HR.
429000 PPH
Fuel 0 MMBTU/HR.
0 KWH
WHRB
Supplementary Fuel 0 MMBTU/HR.
0 PPH
Fuel 0 MMBTU/HR.
0 KWH
WHRB
Supplementary Fuel 0 MMBTU/HR.
0 PPH
Feed Water @ 228°F

FIGURE 11
DECOUPLING NEED FOR STEAM FROM DISPATCHERS NEED TO CONTROL ELECTRICAL OUTPUT

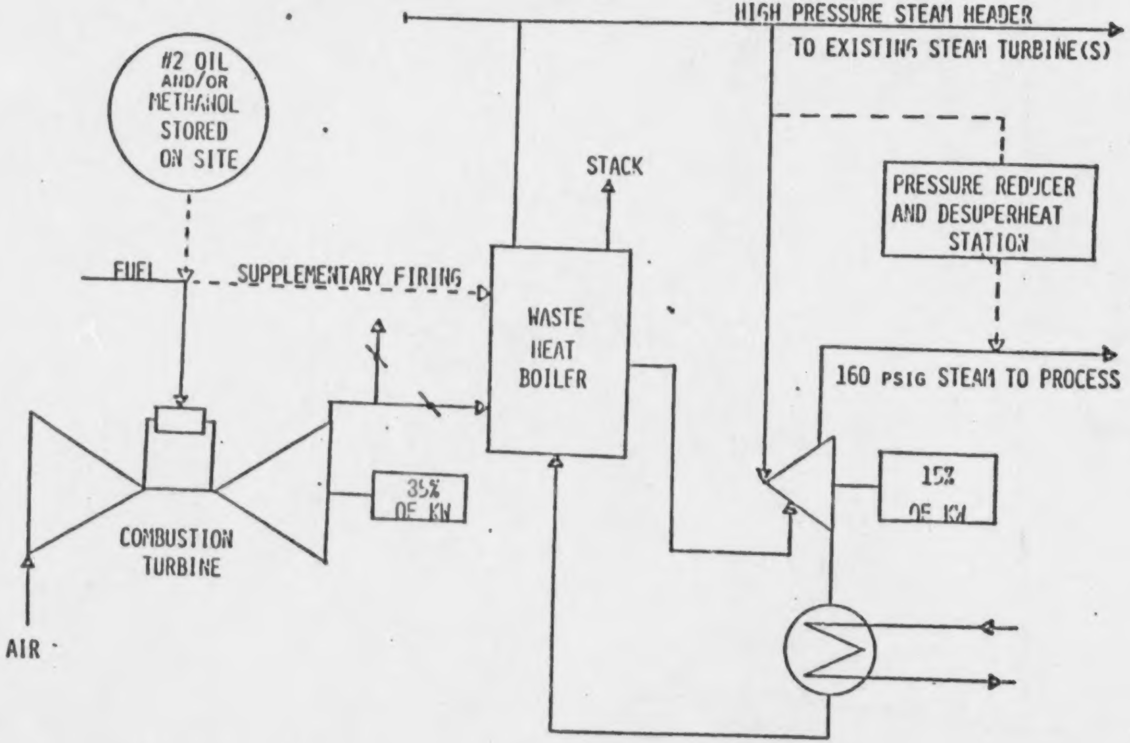
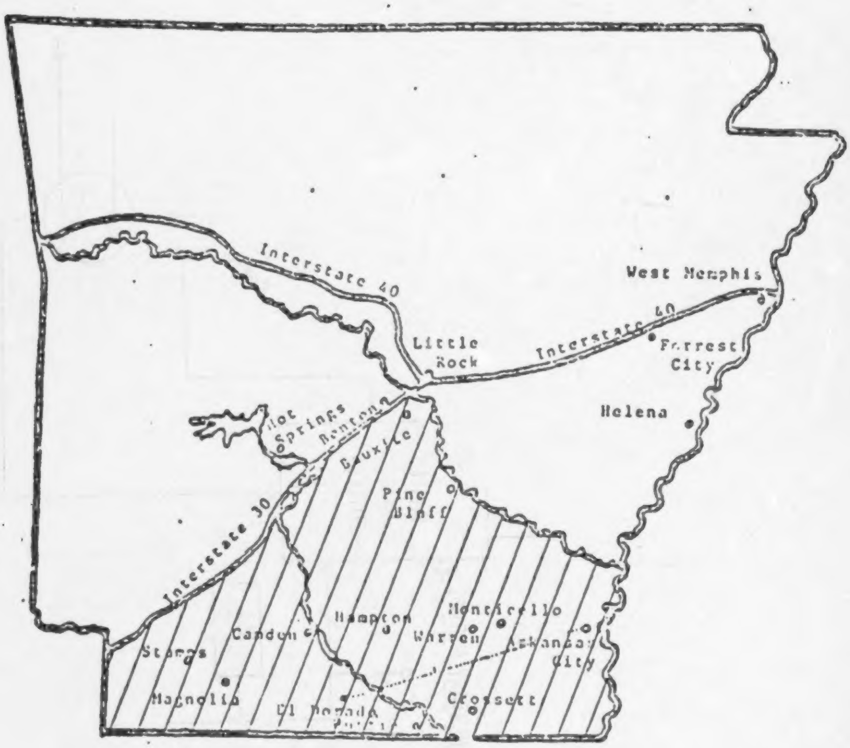

HIGH PRESSURE STEAM HEADER
TO EXISTING STEAM TURBINE(S)
#2 OIL AND/OR METHANOL STORED ON SITE
STACK
PRESSURE REDUCER AND DESUPERHEAT STATION
FUEL
SUPPLEMENTARY FIRING
WASTE HEAT BOILER
160 PSIG STEAM TO PROCESS
35% OF KW
15% OF KW
COMBUSTION TURBINE
AIR

FIGURE 12


Interstate 40
West Memphis
Little Rock
Interstate 40
Forrest City
Helena
Hot Springs
Benton
Bauxite
Pine Bluff
Interstate 30
Monticello
Hampton
Camden
Warren
Arkansas City
Stamps
Magnolia
El Dorado
Crossett

APPENDIX A

The superior fuel savings capability of combined cycle cogeneration -

As compared with:

- Conventional cogeneration technology and
- Purchased electricity and production of steam on-site in steam only boiler installations.

APPENDIX A

Cogeneration is an alternate means of producing manufactured forms of energy. Some of the forms of manufactured energy which can be cogenerated are:

- o Electricity
- o Steam
- o Hot and Chilled Water
- o Hot and Chilled Air
- o Shaft Horsepower

The value of cogeneration lies in our ability to structure such a system so it consumes less fuel than had each form of cogenerated energy been produced in separate fuel-to-energy conversion facilities in the conventional manner.

The thermal forms of manufactured energy can be produced in their separate fuel-to-energy conversion equipment at relatively high efficiency. The percent efficiencies for steam only boilers run from the low 70's for moisture laden fuels such as green wood to the mid and upper 80's for coal and oil. The percent efficiency for producing electricity and shaft horsepower range from the mid 20's to the mid 30's. Some 60% of the energy in the fuel used to produce shaft horsepower and electricity is rejected to the environment at temperature levels too low to be useful.

The fuel saving capabilities of various cogeneration configurations lie in structuring the cogeneration system so that the heat rejected during the process of producing electricity and/or shaft horsepower is rejected at temperature levels which can be used in the production of thermal forms of energy needed in various processes. For this reason, shaft horsepower and/or electricity must be one of the cogenerated forms of manufactured energy produced by a cogeneration plant.

Manufactured forms of energy cannot be effectively stored and, with the exception of electricity, cannot be cost effectively transmitted any reasonable distance. For these reasons, manufactured energy, again with the exception of electricity, must be consumed near where it is produced, and there is a limited geographical area in which it can be marketed. Because electricity can be cost effectively transmitted great distances, it can and is marketed over large geographical areas. For this reason, its production rather than shaft horsepower is included in most cogeneration facilities. With traditional technology, electricity is produced by converting shaft horsepower into electric energy. Developing technologies which can be included in cogeneration installations such as the fuel cell make possible the production of electricity separate, rather than sequential to the production to shaft horsepower.

With the use of conventional cogeneration technologies (See Figure 1-A) the amount of electricity produced usually approximates the electric requirements of the manufacturing facilities which require the steam from which the electric energy is cogenerated. Under these conditions, the electricity can be consumed on site and there is no need to be concerned about the marketing of electricity. When the more advanced cogeneration technology, such as the combined cycle cogeneration (C^3) plants (See Figures 2-A and 3-A) is used, the potential fuel savings are significantly increased. This increase occurs because this technology requires less process steam to produce a kilowatt (3-5 lbs. per kilowatt) than conventional technology (25-50 lbs. per kilowatt). The C^3 technology produced 8-10 times as many cogenerated kilowatt hours as conventional technology. The C^3 approach generally also produces 8-10 times the electricity which can be used by the manufacturing operation consuming the steam from which the electricity is cogenerated. Under these conditions, the means of marketing the cogenerated kilowatt hours is required if the potential fuel saving economies of the C^3 approach is to be realized.

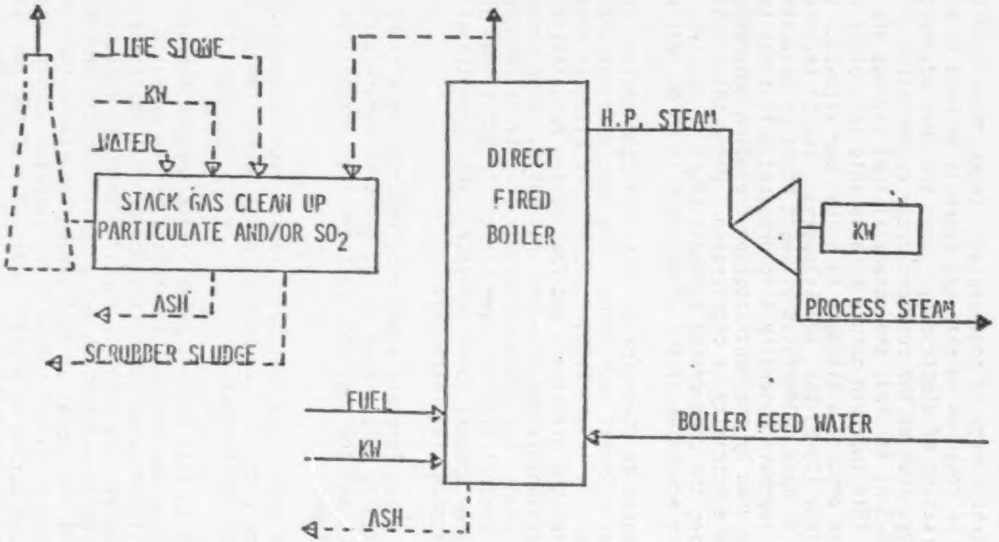
The fuel chargeable to power for conventional cogeneration is in the 4,000-5,000 BTU's per kilowatt hour range. The heat rate chargeable to power for the C^3 plant is in the 5,000-6,000 BTU per kilowatt range. Traditionally, we view the efficiency and fuel saving capabilities of fuel-to-energy conversion systems on a per unit basis. When conventional cogeneration and C^3 plants are compared in this traditional manner to the heat rate of modern central station power plants (10,000-11,000 BTU/KWH), conventional technology appears to be superior. This traditional view fails to take into account the following:

- The process steam (process heat) requirements are a fixed finite quantity which is established by the needs of manufacturing processes, etc., and the cogeneration technology has no impact on their magnitude.
- The C^3 plant saves less fuel per kilowatt hour cogenerated but produces some 8 to 10 times more kilowatt hours from the same process steam requirement.

The improvement in fuel savings ca[illegible] through the use of the C^3 approach rather than conventional [illegible] be dramatic. These savings remain significant when the e[illegible]al gasification equipment producing medium BTU gas (MBG) is [illegible] its fuel saving characteristics. Figures 4-A through 9-A [illegible]dix present a generic comparison of the amount of fuel [illegible]able through the use of conventional cogeneration and the C^3 [illegible]oach. This comparison is somewhat simplistic but it is a valid indication of the fuel savings available through the use of these technologies. The fuel saving for Case IV, which is set out in Figure 9-A, can be somewhat misleading unless consideration is given to the fact that the fuel input to the abbreviated C^3 configuration also supplies the fuel for the electricity generated in the steam turbines of the existing cogeneration plant being repowered. For the purpose of considering the amount of fuel saved, Cases I, III, and IV should be compared. Case II cogeneration produces a fuel reduction over Case I of some 10% for either coal, oil, or natural gas. Case III produces fuel reductions of 35% for natural gas. While the performance of the C^3 plant on #2 oil is somewhat different than for natural gas, it is not significantly different. Thus, the 35% figure is applicable for comparing Cases I & II on #2 oil with Case III. When a coal gasification efficiency is factored into the Case III performance, the fuel saving capabilities comparing coal for Cases I & III is some 24%. The prospect of reducing our consumption of coal 24% below the Case I performance and 14% below the Case II performance is worthy of diligent effort to establish the gasification-C^3 system to supply potential cogeneratable forms of manufactured energy consumed in stationary facilities in this nation.

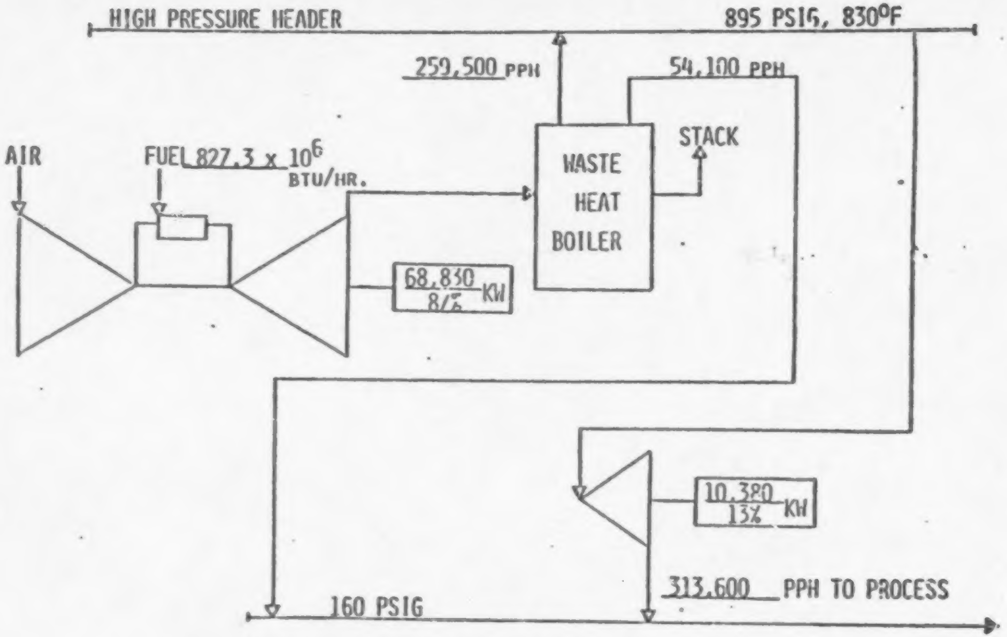
FIGURE 1-A

CONVENTIONAL DIRECT COMBUSTION COGENERATION


LIME STONE
KW
WATER
STACK GAS CLEAN UP PARTICULATE AND/OR SO_2
ASH
SCRUBBER SLUDGE
DIRECT FIRED BOILER
H.P. STEAM
KW
PROCESS STEAM
FUEL
KW
ASH
BOILER FEED WATER

1. LBS. OF STEAM TO PRODUCE ONE KWH (DEPENDS ON DELTA STEAM CONDITIONS OF PROCESS STEAM AND H.P. STEAM AT TURBINE THROTTLE). = . 25 - 50 LBS. PER HR.
2. HEAT CHARGEABLE TO ELECT POWER . = 4000 - 4000 BTU/KWH

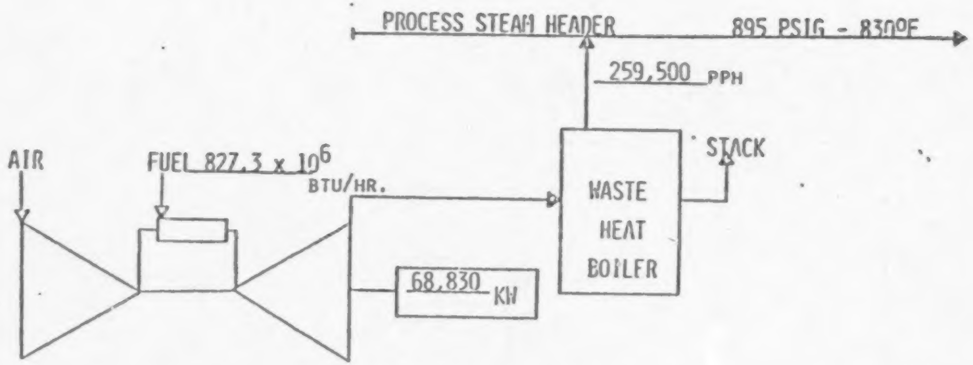
FIGURE 2-A
COMBINED CYCLE COGENERATION
GENERAL ELECTRIC MS7001 (E) NATURAL GAS


HIGH PRESSURE HEADER
895 PSIG, 830°F
259,500 PPH
54,100 PPH
STACK
AIR
FUEL 827.3 x 10^6 BTU/HR.
WASTE HEAT BOILER
68,830 KW
87%
10,380 KW
13%
160 PSIG
313,600 PPH TO PROCESS

1. LBS OF PROCESS STEAM TO PRODUCE ONE KW = 4.0 LBS/KW
2. FUEL TO PRODUCE ONE KW AND 4.0 LBS OF PROCESS STEAM = 10,400 BTU/KWH
3. ELECTRICAL OUTPUT (GROSS) = 79,210 KW

(NOTE: DATA FROM GE BULLETIN GER 2229H).

FIGURE 3-A
COMBINED CYCLE COGENERATION
GENERAL ELECTRIC MS7001 (E) NATURAL GAS


PROCESS STEAM HEADER
895 PSIG - 830°F
259,500 PPH
STACK
AIR
FUEL 827.3 x 10^6 BTU/HR.
WASTE HEAT BOILER
68,830 KW

1. LBS OF PROCESS STEAM TO PRODUCE ONE KWH = 3.8 LBS/KWH
2. FUEL TO PRODUCE ONE KWH AND 3.8 LBS OF PROCESS STEAM = 12,020 BTU/KWH
3. ELECTRICAL OUTPUT (GROSS) = 68,830 KW

(NOTE: DATA FROM G.E. BULLETIN GER-2229N)

FIGURE 4-A

CASE I

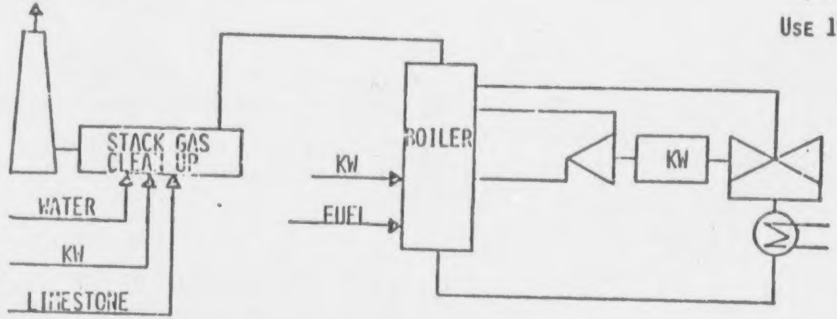
PURCHASED ELECTRICITY & PROCESS STEAM PRODUCED ON SITE

A. PURCHASED ELECTRICITY

HEAT RATE

10,000 - 11,500

USE 10,700 BTU/KWH


STACK GAS CLEAN UP
WATER
KW
LIMESTONE
KW
FUEL
BOILER
KW

B. PROCESS STEAM GENERATED ON SITE (NO STACK GAS CLEAN UP)

HEAT RATE (FUEL INPUT PER LB. OF STEAM)

1200 - 1500

USE 1350 BTU/LB.

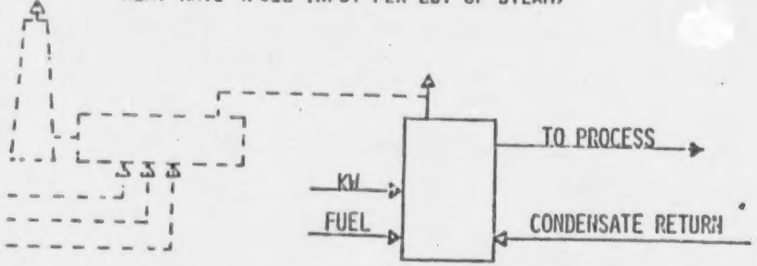

TO PROCESS
KW
FUEL
CONDENSATE RETURN

FIGURE 5-A

CASE II

CONVENTIONAL COGENERATION SYSTEM

C. DIRECT COMBUSTION STEAM TOPPING SYSTEM SUPPLYING 100,000 PPH AT 160 PSIG D & S AT 80% ANNUAL LOAD FACTOR

1. PROCESS STEAM TO POWER RATIO (25 - 50 LBS/KW DEPENDING ON PRESSURE DIFFERENTIAL BETWEEN STEAM TURBINE THROTTLE AND PROCESS STEAM. USE 38 LBS/KW
2. HEAT CHARGED TO PROCESS STEAM 1350 BTU/LB
3. HEAT CHARGED TO ELECTRIC POWER 4200 - 5000 USE 4500 BTU/KWH
4. ANNUAL FUEL SAVINGS
    - A. FUEL SAVING FOR COGENERATED STEAM (100,000 x 1350) - (100,000 x 1350) 0 BTU
    - B. FUEL SAVING FOR COGENERATED ELECTRICITY (100,000 ÷ 38) x (10,700 - 4500) x (8760 x .8) 114.4×10^9 BTU/YR

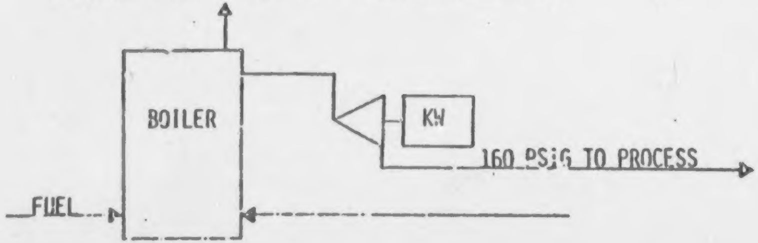
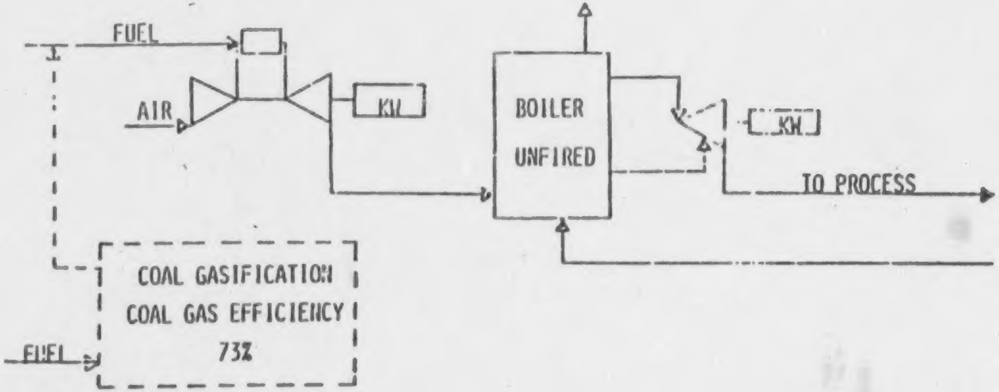


FIGURE 6-A

CASE III
COMBINED CYCLE COGENERATION SYSTEM

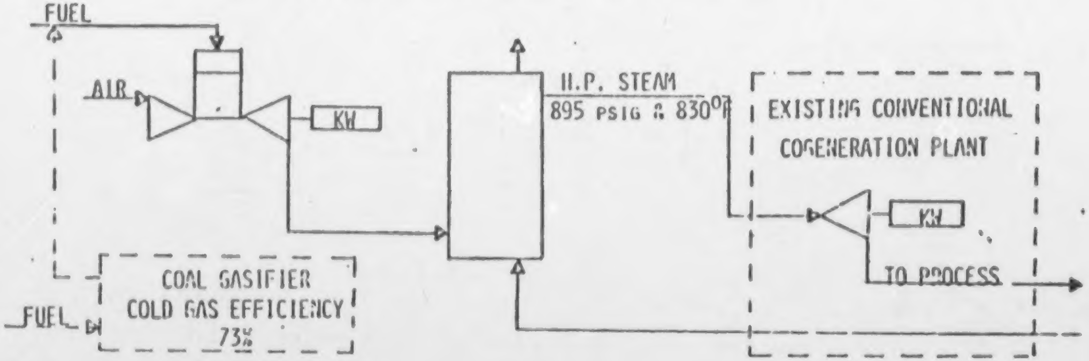
D. Combined cycle - combustion and steam turbine supplying 100,000 pph at 160 psig D & S at 80% annual load factor

1. Process steam to power ratio - 3 - 5 lbs/kw — Use 4 lbs/kw
2. Fuel input to produce 4 lbs steam & 1 kw — 10,400 btu/hr
3. Fuel chargable to power 10,400 - (4 x 1350) — 5,000 btu/kwh
4. Annual fuel saving using natural gas (for steam)
    - a. (100,000 ÷ 4) x (10,700 - 5,000) x (8760 x .8) = 998.6 x 10^9 btu/yr
5. Annual fuel savings using mbg from coal gasifier
    - a. 25,000 kw x ((10,700 - (5000 ÷ .73)) x 7008 hrs — 674.7 x 10^9 btu/yr


FUEL
AIR
KW
BOILER
UNFIRED
KW
TO PROCESS
COAL GASIFICATION
COAL GAS EFFICIENCY
73%
FUEL

CASE IV
COMBINED CYCLE COGENERATION FOR REPOWERING EXISTING CONVENTIONAL COGENERATION

FIGURE 7-A


FUEL
AIR
KW
H.P. STEAM
895 PSIG & 830°F
EXISTING CONVENTIONAL
COGENERATION PLANT
KW
TO PROCESS
COAL GASIFIER
COLD GAS EFFICIENCY
73%
FUEL

D. COMBINED CYCLE IN ABBREVIATED CONFIGURATION SUPPLYING 100,000 PPH AT 895 PSIG, 830°F AT 80% ANNUAL LOAD FACTOR

1. PROCESS STEAM TO POWER RATIO 3.81 LBS/KW
2. FUEL INPUT TO PRODUCE 3.8 LBS. STEAM & 1KW 12.020 BTU
3. FUEL CHARGABLE TO POWER 12,020 - (1350 x 3.8) = 6620 BTU/KWH
4. ANNUAL FUEL SAVING USING NATURAL GAS
   A. (100,000 ÷ 3.8) x (10,700 - 6,620) x (8,760 x .8) = 752.4 x 10^9 BTU/YR.
5. ANNUAL FUEL SAVINGS USING MBG FOR GASIFIER WITH COLD GAS EFFICIENCY OF 73%
   A. 26,316 KW x ((10,700 - (6,620 ÷ .73) x 7008 HRS. = 301.0 x 10^9 BTU/YR.

FIGURE 8-A

ESTIMATED FUEL SAVING CAPABILITY OF COGENERATION PER 100,000 PPH OF PROCESS STEAM

	Cogenerated Elect. Cap. KW	Process Steam PPH	Annual Fuel Requirement at 80% Load Factor				
			100,000 PPH Steam 10^9 Btu	Cogenerated Electricity 10^9 Btu	Total Col. 3 & Col. 4 10^9 Btu	BBL #6 Oil at 6.3 x 10^6 Btu/bbl	Tons of Coal at 11,000 Btu/lb.
Case II							
Case I Equivalent	[illegible],632	100,000	946.08	197.36	1,143.44	181,498.41	51,975
Case II	2,632	100,000	946.08	83.00	1,029.03	163,346.03	46,776
Fuel Saved			0	114.36	114.36	18,152.38	5,198
% Reduction					10%		
Case III							
Case I Equivalent	25,000	100,000	946.08	1,874.64	2,820.72	447,733.33	128,215
Case III Natural Gas	25,000	100,000	946.08	876.11	1,822.19	289,236.51	62,827
Fuel Saved				998.53	998.53	158,496.83	45,388
% Reduction					35.4%		
Case III MBG	25,000	100,000	946.8	1,200.00	2,146.08	340,647.62	97,549
Fuel Saved				674.64	674.64	107,085.71	30,665
% Reduction					24%		
Case IV							
Case I Equivalent	26,316	100,000	946.08	1,973.32	2,919.40	463,396.83	132,700
Case IV Natural Gas	26,316	100,000	946.08	1,220.88	2,166.96	343,961.90	98,498
Fuel Saved				752.44	752.44	119,434.92	34,202
% Reduction					25.77%		
Case IV MBG	26,316	100,000	946.08	1,672.53	2,618.61	415,652.38	119,028
Fuel Saved				300.79	300.79	44,744.44	13,672
% Reduction					10.3%		

FIGURE 9-A

FUEL SAVING IMPROVEMENT THROUGH THE USE OF COMBINED CYCLE COGENERATION

	ELECTRIC COGENERATED CAPACITY KW PER 100,000 PPH OF PROCESS STEAM	INCREASED COGENERATED ELECTRIC CAPACITY %	FUEL SAVED BY USING COGENERATION 10^9 BTU/YR	INCREASED FUEL SAVING %
CASE II				
CONVENTIONAL COGENERATION	2,632	BASE	114.4	BASE
CASE III				
COMBINED CYCLE COGENERATION				
A. NATURAL GAS	25,000	850	998.6	773 } compared to
B. MBG	25,000	850	674.7	490
CASE IV				
COMBINED CYCLE COGENERATION IN REPOWERING CONFIGURATION				
A. NATURAL GAS	26,316	900	752.4	558
B. MBG	26,316	900	301.0	163

The Sulf-X process is a wet, regenerative scrubbing system that has demonstrated high SO2 and NOx removal capability. This report from Ellison Consultants reviews the demonstration project at Canonsburg, Pennsylvania, and examines the technical feasibility and economics of the process if applied to a 500 power plant.

THE SULF-X FLUE GAS DESULFURIZATION PROCESS

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Presentation at:

Tenth Energy Technology Conference and Exposition, (Session D-1, Burning Coal Cleanly,) Sheraton Washington Hotel, Washington, D.C., February 28, 1983

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ABSTRACT

The newly developed Sulf-X Process, a regenerative wet-scrubbing technique, shows great promise as a practical and economical means for efficient collection of sulfur oxides emissions from large coal-fired boilers. SO_2 in boiler flue gas reacts with a slurry of ferrous sulfide that is thereafter regenerated in an indirect coal-fired calciner, which yields an elemental-sulfur off-gas that is condensed and thereby made available as a by-product for commercial use.

A description is given of a stoker-boiler-size Sulf-X demonstration installation completed in 1982 at the coal-fired boiler plant of the Commonwealth of Pennsylvania's Western Center Hospital in Canonsburg, Pennsylvania, and of startup work and operating experience. Data from program activities to date characterizes gas cleaning performance in an absorption-reduction scrubbing mode in which sulfur dioxide and nitrogen oxides are simultaneously removed from the raw flue gas stream.

Technical feasibility and economic attractiveness of the process are examined at a 500 megawatt power plant scale; and contemplated scale-up of the Western Center operations is reviewed.

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I. INTRODUCTION

The Sulf-X Process is a wet absorption process that utilizes a slurry of continuously regenerated ferrous sulfide solids to provide highly efficient scrubbing of combustion gases to remove a minimum of 90 to 98% of sulfur dioxide gases. A coal-fired sulfide-regeneration system converts collected sulfur dioxide assimilated by sulfides into salable elemental sulfur by-product with minimal formation of non-regenerable waste solids and liquids. The process is presently in small commercial scale demonstration and testing that follows from six years of laboratory, bench-scale, and prototype system testing and development. Ongoing proving of the process is strongly encouraged by the comparative simplicity of the means for production of valuable elemental sulfur by-product, the low cost of ownership and operation in comparison with other regenerative processes, and the 1980s trend toward such open-loop FGD systems brought about by commercial deployment of forced-oxidation operation in lieu of natural-oxidation mode operation in limestone scrubbing applications.

II. PROCESS DESCRIPTION

Ideally the Sulf-X Process is used after dry fly-ash collection, as well as pre-scrubbing when required for segregated low-pH absorption of tramp constituents such as hydrogen chloride. In the FGD operation, schematically illustrated in Figure 1, an aqueous slurry of iron/sulfur compounds, primarily finely divided ferrous sulfide (FeS) reagent in excess along with secondary iron and sulfur species, is used to absorb sulfur dioxide (SO_2) in a wet scrubber. By providing a chemically reducing (absorption-reduction) environment in the liquid phase, and with provision for extended gas residence time in the scrubber, a major proportion of nitrogen oxides present may also be collected. Suspended solids in the bleed from the scrubber are dewatered by thickening and centrifuging and thereafter

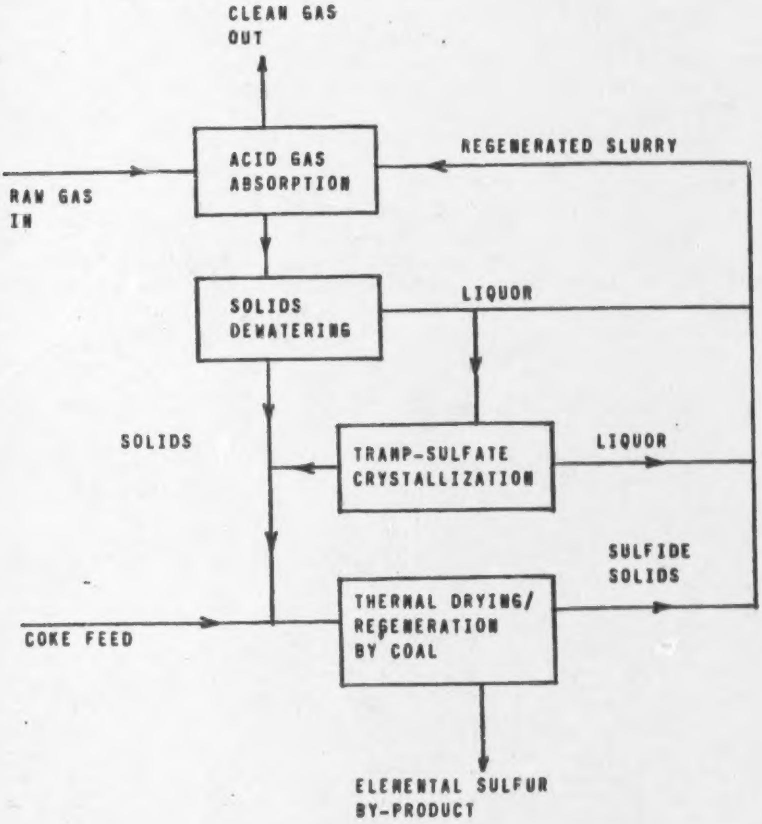
CLEAN GAS OUT
ACID GAS ABSORPTION
REGENERATED SLURRY
RAW GAS IN
SOLIDS DEWATERING
LIQUOR
SOLIDS
TRAMP-SULFATE CRYSTALLIZATION
LIQUOR
SULFIDE SOLIDS
THERMAL DRYING/ REGENERATION BY COAL
COKE FEED
ELEMENTAL SULFUR BY-PRODUCT

Figure 1. Sulf-X Process Flow Diagram

thermally dried and then roasted in an indirect-fired calciner using coal fuel. The thickener overflow stream, principally containing dissolved sodium sulfate (Na_2SO_4) and ferrous sulfate ($FeSO_4$), is discharged to a calciner quench tank to both cool and slurry the hot solids leaving the calciner. However, a portion of the overflow stream is diverted to a crystallizer system to form sodium sulfate solids. Net loss of sodium, if any, from the system is made up by addition of sodium sulfide. After the dewatered crystallizer product is combined with the absorber-product centrifuge cake for drying, a proportion of coke is added at the calciner inlet. In the calciner at 1400°F, sulfur-rich ferrous sulfide formed in the scrubber dissociates to ferrous sulfide and elemental sulfur, and the sodium sulfate is reduced to sodium sulfide. The sodium sulfide subsequently reacts with dissolved ferrous sulfate in the quench tank to precipitate ferrous sulfide. Sulfur released during the calcination leaves the calciner as a vapor and is recovered by condensation. By-product steam is generated in the sulfur condensation step as well as in an unfired waste heat boiler applied to the calciner exhaust combustion gases. The Sulf-X Process sulfate-make is crystallized and thermally reduced in the form of sodium sulfate because reduction to sulfide is thereby accomplished without liberation of SO_2 that would accompany calcining of alternative sulfate forms such as ferrous sulfate. Calcination of sodium sulfate is a common commercial process. Moreover, in the Sulf-X Process, iron compounds that are present catalyze the reduction of sodium sulfate, and the minimum temperature for its reduction is less than the design temperature for the concurrent calcination of the sulfur-rich FeS forms.

III. PRINCIPAL CHEMICAL REACTIONS

A. ABSORPTION REACTIONS, SO_2 REMOVAL ONLY: When the system is engineered at a typical FGD-process superficial gas velocity of approximately 10 feet per second, minimal NO_x removal occurs. The overall chemical reaction describing SO_2 absorption in the scrubber is complex, but may be expressed algebraically as:

$$(c)FeS+SO_2+O_2 \xrightarrow{H_2O} (c-1)FeS_{(m)} + Fe^{++} + SO_4^{=} + \left(1-\frac{n}{10}\right)S^0$$

Intermediate absorption reaction products include HS^-, $SO_3^=$, HSO_3^-, and $S_2O_3^=$ (thiosulfate). Secondary reactions not reflected include oxidation of ferrous ions to ferric, and oxidation of bisulfite and sulfite ions to sulfate. The quantity "n" above, an integer that can range from 1 to 10, is a measure of the degree of sulfidation of the reagent arising from its direct reaction with elemental sulfur formed in the absorber, and reflects the spectrum of reaction products between FeS(fresh) and FeS_2(fully spent). The quantity "(c)" is greater than 1 but less than 2, its magnitude varying from application to application, and from one operating condition to another, depending on redox (reduction-oxidation) equilibrium. In commercial design, FeS is typically fed to the absorption reaction based on the design assumption that c=2 moles FeS per mole SO_2 absorbed. The quantity "(m)" equals (n/10+c-1)/(c-1). Analyses have shown that the particles of SO_2-reaction-product consist of elemental sulfur and sulfur-rich iron sulfides (FeS_m), with unreacted FeS at the core. Studies have shown that neither x-ray diffraction nor chemical analyses of these type materials are sufficiently selective to fully and accurately define the distribution of the iron/sulfur compounds formed. However, x-ray and bulk analyses have qualitatively confirmed the presence of iron sulfides and elemental sulfur, the latter having been extracted with carbon disulfide under laboratory conditions.

B. ABSORPTION REACTIONS, SIMULTANEOUS SO_2/NO_x REMOVAL: When the scrubbing is designed for catalysis by a significant concentration of ferrous ions, e.g. 40 grams/liter, and with flue-gas residence time in the scrubber sufficient to achieve appreciable NO_x removal, the overall chemical reaction mechanism becomes:

$$(c)FeS+SO_2+NO+\tfrac{1}{2}O_2 \xrightarrow[H_2O]{Fe^{++}} (c-1)FeS_{(m)}+Fe^{++}+SO_4^{=}+ (1-\tfrac{n}{10})S^{o}+\tfrac{1}{2}N_2\uparrow$$

Additional intermediate reaction products include $Fe(NO)^{++}$.

C. REGENERATION REACTIONS: Individual chemical reactions in the regeneration system are as follows:

$$FeS_{1+\frac{n}{10}} \xrightarrow{\Delta} FeS + \tfrac{n}{10}\,S^{o}\uparrow \quad . \qquad \text{(Calciner)}$$

$$Na_2SO_4 + 2C \xrightarrow{\Delta} Na_2S + 2CO_2\uparrow \qquad \text{(Calciner)}$$

$$Na_2S+FeSO_4+H_2O \longrightarrow FeS\downarrow +Na_2SO_4+H_2O \qquad \text{(Quench Tank)}$$

IV. PROCESS ADVANTAGES

Due to the simplicity of design and economy of operation, the annual revenue requirements for Sulf-X application are expected to be lower than other available methods for high-efficiency acid-gas cleanup:

Regenerable ferrous iron compounds are used as acid-gas absorbent in lieu of non-recoverable lime or limestone alkalis.

Salable sulfur is produced rather than SO_2 sludge.

Tramp sulfate-make is continuously regenerated so as to avoid need to purge any sulfate waste product. Tramp dithionate production does not occur.

Minimal iron losses are replenished through makeup by inexpensive industrial by-product iron compounds such as pyrites, (FeS_2).

Thermal regeneration utilizes coke and boiler coal-fuel, and the process requires no premium fuels or raw material derivatives thereof such as ammonia or reducing gases.

The salable sulfur yield is a direct product of the coal-fired regeneration step and thus no further chemical processing of this regenerator by-product is needed.

When designed with provision for enhanced gas-residence time in the wet scrubbers, application of the process for flue gas desulfurization provides an absorption-reduction chemical environment , and from 60 to 90% of nitrogen oxide gases are simultaneously removed and converted to innocuous elemental nitrogen (without formation of imidodisulfonate or other nitrogen/sulfur tramp ion complexes typical of EDTA-catalysed absorption-reduction processes.)

V. PROCESS DEVELOPMENT, TESTING, AND DEMONSTRATION

Development of the Sulf-X Process through laboratory and field testwork was completed in 1980. Design and construction of a stoker-boiler size integrated process installation for the Commonwealth of Pennsylvania was completed in 1982. This demonstration facility is now in operation and will be used for process scale-up in pulverized coal-fired boiler applications.

A. PRELIMINARY PROCESS STUDIES: Beginning in 1976 an intensive laboratory, field pilot-plant, and bench-scale test program was carried out to establish the present Sulf-X Process design. After earliest laboratory studies of the gas absorption characteristics and thermal regeneration of ferrous-sulfide type compounds, a pilot-plant scrubber using this Sulf-X reagent was field tested on boilers firing a diversity of bituminous coal fuels, and demonstrated SO_2 removal efficiency as high as 96% with NO_x removal levels ranging to higher than 60%. Gas absorption mechanisms were subsequently studied in the laboratory under the sponsorship of the Department of Energy (D.O.E.) to characterize the catalysis of NO_x removal by ferrous ion in the presence of suspended ferrous sulfide solids. D.O.E. commissioned an advanced bench-scale technical development program, completed in 1980, which utilized a bench-scale replica of the entire process, absorption as well as regeneration. It assessed integrated closed-loop process operation treating actual coal-fired boiler emissions containing adversely high oxygen content, e.g. ranging in excess of 10% by volume. Accompanied by SO_2 removal at the 95% level with NO_x removal above 80%, ferrous sulfide reagent successfully underwent more than 10 repetitive cycles of regeneration. It was determined in subsequent studies that suitable calcination/recovery of sulfate production, the principal tramp product of parasitic oxidation effects in the process, requires that the sulfate be precipitated/crystallized as sodium sulfate for feed to the thermal calcination step. Tests have shown that with the typically high concentration of iron compounds present in Sulf-X Process operation, a rapid chemical reduction of the sodium-sulfate intermediate takes place at the same temperature used for calcining the sulfur-rich spent ferrous sulfide solids.

B. PURPOSE OF EXISTING PROCESS DEMONSTRATION FACILITIES: An integrated Sulf-X Process completed in 1982 is now in demonstration operation at the bituminous coal-fired boiler plant of the Western Center Hospital of the Commonwealth of Pennsylvania. The system is designed for the simultaneous control of SO_2 and NO_x emissions from one of the plant boilers at a flue gas flow rate of approximately 7000 acfm, with capability for removal of SO_2-only for flue-gas throughput in excess of 10,000 acfm. Its current test/demonstration operation in 1983 under funding by the Commonwealth of Pennsylvania constitutes a critical interim stage in the application of the process through collection of steady-state operating data in anticipation of scale-up to a later large commercial installation. The continuous operation of this demonstration facility at the same time affords a needed opportunity to study the effects on the overall chemistry of reaction products and tramp materials that accumulate in the system inventory, and to evaluate the compatability of the equipment and instrumentation installed.

C. DESCRIPTION OF DEMONSTRATION FACILITIES: The Sulf-X installation at Canonsburg is a $2 million facility dedicated by the Commonwealth of Pennsylvania to an initial 12-month period of demonstration activity in conjunction with characterization and optimization of process operation. It advantageously utilizes off-the-shelf process equipment and control instrumentation to provide a practical replica of the process facilities that are expected to be later applied in large commercial installations. Thus, the system models a large-scale installation by duplicating the primary unit operations such as gas absorption, slurry dewatering and thermal/chemical regeneration, using applicable mass transfer devices and mechanical equipment. Figure 2 shows the scrubber tower at Canonsburg and the Sulf-X Process building adjoining the boiler house at the left. Flue gas cleaning equipment operating includes a 3-feet diameter radial-blade booster fan, a lined carbon-steel gas precooler, and a 73 feet high by 5-feet diameter FRP scrubber (absorber) vessel using cross-fluted fixed plastic packing. Slurry-bleed dewatering equipment consists of a 10-feet diameter lined thickener with overflow tank, a horizontal plate and frame filter containing 17 1-meter x 1-meter vertical filter plates, and a steam-heated rotary dryer (Figure 3.) The regeneration system consists of an indirect gas-fired rotary calciner (Fig.4), a calcine quench tank, and a steam-cooled vertical shell-and-tube sulfur condenser. For supply of regenerated reagent to the gas absorption system, the system includes an agitated-media type calcine-slurry crusher (attritor) and an absorber fresh-slurry feed tank.

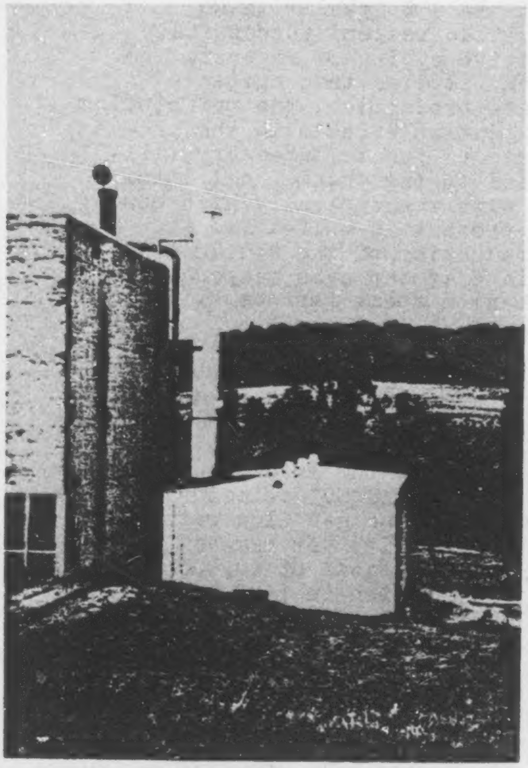

Figure 2. Sulf-X Demonstration

To simplify the overall demonstration program without adversely affecting the simulation of operation of large commercial systems, sodium sulfate crystallization facilities have been excluded. Accordingly, instead, dissolved sodium sulfate is purged from the system by blowdown, and purchased sodium sulfide is used (in lieu of sodium sulfide that would otherwise be available from calcination of sodium sulfate) to sustain the quench tank reaction, which recovers dissolved ferrous sulfate by converting it to precipitated FeS reagent. To further simplify the operation of this small-capacity demonstration operation, the calciner is fired with LPG (propane) instead of coal. Since calcination is by

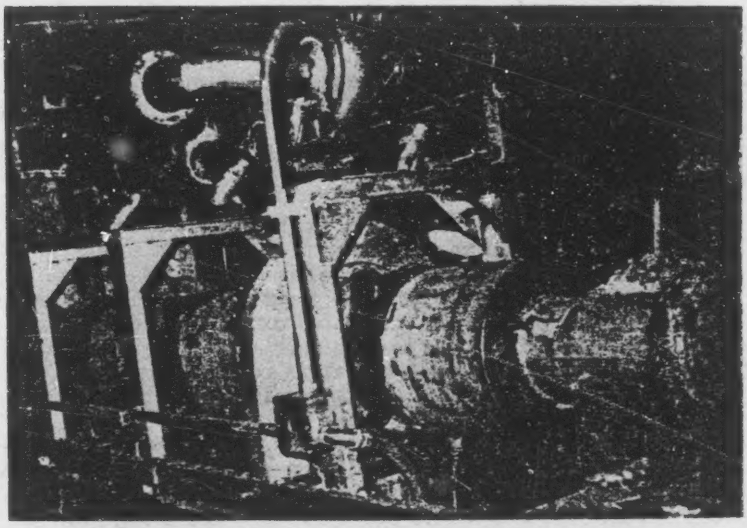

Figure 3. Steam-Heated Rotary Dryer

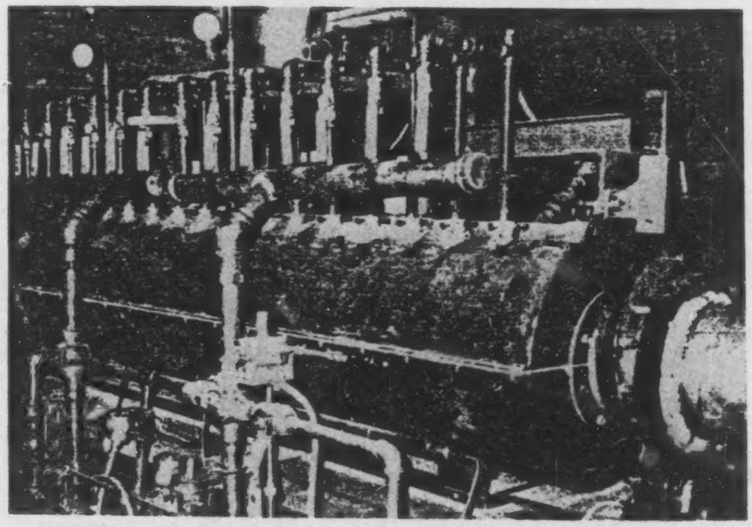

Figure 4. Indirect Fired Rotary Calciner

indirect heat transfer both in the commercial scale and Western Center designs, use of alternative fuels does not effect process chemistry or the validity of the test operations. In addition, pre-cooling and water saturation of the boiler flue gas is carried out in a simple quenching step utilizing recirculating process slurry. In absence of an isolated pre-scrubbing wastewater purge, chloride will be purged along with the dissolved sodium sulfate blowdown liquid. As a result of other simplications, calciner combustion gases are exhausted without by-product steam generation, and sulfur-condenser off-gas is recycled to the absorber inlet instead of being treated in an incinerator.

D. STARTUP PROBLEMS EXPERIENCED: Principal startup/operating problems encountered with the initial system installation are difficulties in mechanically transferring the filter cake to the dryer and in moving the filter cake through the drying operation; and the oxidation effects of excessive amounts of atmospheric air entering the process-side of the calciner. Because of the small batch-scale nature of the filtering operation at Western Center, transport of wet filter cake is expected to continue to require special operator attention during continuous process runs. The original dryer and calciner have been replaced in late 1982 with the more compatible equipment described above, and these operations are now considered to be free of major problems. Retrofit modifications have also been made to the sulfur-vapor vacuum and condensing systems to avoid solidification of sulfur before the condenser as well as carryover of sulfur liquid and vapor to the vacuum producing equipment downstream.

E. PROGRAM OF SYSTEM TESTING: The Western Center scrubber has successfully demonstrated up to 99+% removal of SO_2 and up to 93% removal of NO_x in 2%-sulfur bituminous coal service. At a superficial gas velocity in the scrubber of approximately 10 feet per second, typical of operation of commercial FGD systems, SO_2 removal at an efficiency of 95+% is accompanied by only token NO_x removal. At a superficial velocity no greater than 4 feet per second, NO_x removal efficiency is 90+% with essentially 100% SO_2 removal. The absorption system has operated automatically with the pH controller regulating fresh-slurry addition and spent-slurry removal to maintain a constant, pre-set pH. Pyrites has been used for iron makeup. During 1982 startup and commissioning activity, spent slurry has dewatered well in the thickener and filter, and the FeS precipitation and chemical makeup systems have performed as designed. The Commonwealth of Pennsylvania is expected to authorize and fund continuous round-the-clock operations beginning in the early 1983 winter months. Comprehensive data is being collected for a broad range of operating conditions to provide parametric testing to optimize system design for design NO_x removal levels up to 80%, and to yield a data base to be used for scale-up of the commonly used sub-processes that comprise the Sulf-X Process. Alternative process operation will be tested in runs that will substitute 4%-sulfur coal for boiler firing, and copperas ($FeSO_4 \cdot 7H_2O$)

for iron makeup. In additional runs, purchased sodium sulfate will be fed with petroleum and/or metallurgical coke to the regeneration system to demonstrate the use of sodium-sulfate intermediate product as the process design means for tramp sulfate recovery. NO_x/SO_2 removal efficiency will be correlated to scrubber liquid/gas ratio and ferrous ion concentration during steady-state operation along with superficial gas velocity and gas residence time.

VI. COMMERCIAL-SCALE PROCESS ECONOMICS

The Sulf-X Process is believed to offer a technically sound and economically attractive means for high-capacity regenerative FGD operation. In-house preliminary conceptual design and costing studies to date have been based on a hypothetical 500 MW(e), 3.5%-sulfur coal application. A 1982 estimate for SO_2-removal-only operation indicates a total capital investment of approximately \$60 million, i.e. \$120/gross kW, including indirect costs, contingency, startup allowance, interest during construction, working capital, and royalty fee. Annual costs with zero by-product credit are projected at approximately \$22 million (1982-\$), representing less than 7 mills/kWh, which includes approximately \$4.5 million for raw materials, (primarily pyrites concentrate and coke), \$0.5 million operating labor, \$4.5 million for utilities, (primarily electric power and process steam), \$2.5 million maintenance, \$8.5 million capital charges, and \$1.5 million overhead costs. (Based on anticipated sale of sulfur by-product at a nominal \$100 per ton price, total annual costs would be reduced to approximately \$18 million.) A 1982/3 detailed study of commercial FGD economics at the 4%-sulfur coal, 500 MW(e) level by the Electric Power Research Institute, scheduled to be published in the first half of 1983, is expected to provide an appropriate direct comparison of the economics of all principal FGD processes being studied including Sulf-X. 1979 in-house studies, on a comparable basis, for commercially available combined SO_2/NO_x removal facilities achieving 80% NO_x removal confirm that Sulf-X is substantially less costly than other available wet simultaneous SO_2/NO_x removal processes. Moreover, while capital cost is indicated to be somewhat higher than for combined removal methods that include dry catalytic NO_x control processes, the Sulf-X Process is estimated to have lower annual revenue requirements than all available combined removal methods including those based on dry NO_x control.

This report from the Electric Power Research Institute (EPRI) discusses a recent decision to emphasize turbocharged pressurized fludized bed combustion (PFBC) systems. This change will enable commercialization of a PFBC sooner then originally anticipated. Commercially available systems should be available for small, coal fired power plants by the late 1980's or early 1990's.

EPRI

RFP 3236

PFBC TURBOCHARGED BOILER DESIGN AND ECONOMIC STUDY

ATTACHMENT 1: THE PFBC TURBOCHARGED BOILER

TECHNOLOGY REVIEW, RATIONALE FOR DEVELOPMENT AND R&D PLAN

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INTRODUCTION

The real unit cost of electricity in the United States declined continuously for over five decades. That decline, which ended in the early 1970s, was achieved through the introduction of new technologies, capture of greater economies of scale and the replacement of old, inefficient units with modern, efficient power plants. As part of the effort to produce lower cost electricity, power researchers worked in the 1960s and 1970s to develop a direct coal-fired combined-cycle power plant that would burn coal in a fluidized-bed combustor at elevated pressure (PFBC). The plant would generate electricity in both a gas turbine expanding the hot, pressurized coal combustion products, and in a conventional steam cycle.

Based on substantial PFBC development efforts to date, EPRI proposes to achieve a near-term application of PFBC. EPRI's strategy will encourage the construction of highly-efficient (about 37%) low-risk, first-generation turbocharged boilers in which the gas turbine operates in a self-sustaining, zero net power output (turbocharging) mode. In this configuration, the gas turbine inlet temperature is under 800°F, permitting the use of substantially conventional filters in the gas stream and virtually eliminating gas turbine erosion and corrosion. The turbocharged boiler system is expected to meet new market demands in the late 1980s and early 1990s for smaller, coal fired power plants capable of rapid field erection and startup. Having confidently established that gas turbines can successfully expand coal combustion products in such a low-risk plant, the technology can further evolve to higher gas turbine inlet temperatures in next-generation plants thus achieving optimum combined cycle-conditions.

It is the objective of this paper to develop the logic behind this goal and to outline EPRI's five-year PFBC R&D plan.

STRATEGY SUMMARY

The following are the key strategic elements of EPRI's revised PFBC R&D plan:

- Emphasis changed to development of turbocharged boiler (zero net gas turbine output) from PFBC combined-cycle. Technical risk in hot-gas cleaning, hot-gas ducting, gas turbine components, and system control is considerably reduced. Reliability/availability of gas turbine is predicted to increase substantially.

- Smaller unit size (100-200 MW_e) development will be emphasized. It presently appears feasible to shop-fabricate and barge-transport units in the 100 to 125 MW_e capacity range.

- The turbocharged boiler will meet specific needs of late 1980s and early 1990s; viz, ability to increase the nameplate rating of existing coal-fired units and repower gas- and/or oil-fired units. Shop assembly and rapid field erection of turbocharged boilers in these applications would significantly reduce utility financial risk.

- Ultimate objective remains to develop PFBC/combined-cycle plants with 8500 Btu/kW hr heat rate. These plants are now viewed as second-generation and the PFBC turbocharged boiler is judged a critical, and at the same time economically attractive, step on the development path of this longer range objective. DOE, in their new role, will continue baseline funding for the combined-cycle PFBC plant.

BENEFITS OF PFBC

Pressurized fluidized-bed combustion power plants have been shown to have attractive benefits for the utility industry. All of the benefits listed below will be realized in the ultimate combined-cycle plant, yet the majority of them will be available in first-generation turbocharged boiler plants:

- Combined-cycle plants are inherently more efficient than simple-cycle plants because they operate over a greater differential temperature between heat source and sink. PFBC/CC plants would use both the drop from 1500°F to 800°F in the gas turbine and 1005°F to 100°F in the steam turbine, compared to the conventional 1005°F to 100°F range in modern steam plants. Because of this relatively large available WT and the comparative simplicity of PFBC cycles, PFBC is predicted to have the lowest levelized coal pile-to-busbar energy cost of any developmental coal system[1,2] (see Figure 1). Burns & Roe, under contract to EPRI[3] found the cost of electricity for PFBC to be about 10% less than that achievable from conventional pulverized-coal fired plants with scrubbers. In this study, they compared the Curtiss-Wright, General Electric and American Electric Power/Stal-Laval commercial plant concepts to a reference PC plant using common EPRI Technical Assessment Guide assumptions. It was shown that the GE and AEP/SL plants promise a levelized busbar power cost about 11% less than the conventional plant for both Eastern bituminous (4%S) and Western subbituminous (1/2%S) coals.

- PFBC combined-cycle plants capable of achieving 40% net thermal efficiencies have simpler thermodynamic cycles as compared to other advanced coal-based options. The coal is burned in a fluidized-bed steam generator operated at elevated pressure, the combustion gases are cleaned to remove entrained particulate that would erode the gas turbine, the gases are expanded through the gas turbine and the waste heat then raises steam in the heat recovery steam generator (HRSG). The steam side of the cycle (Rankine components) utilizes conventional technology. Figure 2 shows a simplified cycle schematic of a PFBC combined-cycle plant.

- SO_2 control occurs during the combustion process, eliminating the need for flue gas desulphurization equipment. It has been consistently demonstrated, both at Curtiss-Wright and at the National Coal Board (NCB) in England in 1979 and 1980, that >90% SO_2 capture can be achieved using dolomitic sorbents at Ca:S stochiometries of <1.7.

- Water consumption, vis-a-vis conventional wet-cooled steam plants, is low in PFBC combined-cycle plants. Water use reduction is proportional to the level of the gas turbine participation in the overall plant thermal cycle (i.e., fraction of total power from gas turbine). High participation plants, such as the air-tube cycle, can have up to a 3:1 advantage over conventional steam cycles with wet cooling because condensing requirements decrease proportionally to reduced power contribution from the steam turbine. The low-gas turbine participation (30%) PFBC combined cycle was compared by Yeager[4] to four other technologies using common 1979 EPRI Technical Assessment Guide assumptions. It was shown to be the lowest water-consuming cycle at 490 gal/hr/MW (27% less than baseline PC).

- Burning fossil fuels at elevated pressure decreases the size of all steam generator components as demonstrated by Foster Wheeler in their investigation of use of heavy-oil-fired turbocharged boilers.[5] By virtue of the increased combustion gas density, total heat transfer surface requirements were reduced by 50% from the standard non-pressure-fired boiler. This reduction occurred even without capitalizing upon the improved heat transfer characteristics of heat transfer surface immersed in the fluid bed, as PFBC plants are capable of doing.

- PFBC has a high inherent level of fuel flexibility. The NCB PFBC, which ran for 1000 hours on high-sulfur Ohio bituminous coal, has also operated very successfully on low-quality coals supplied by the British Columbia Hydro and Power Authority (Hat Creek subbituminous coal; 45% ash, 8% moisture and 0.8% sulfur). The inherent fuel flexibility of the fluidized-bed combustion process has also been demonstrated by Babcock & Wilcox at atmospheric pressure in the Alliance, Ohio 6' x 6' test facility. Successful on-line transfers between Eastern high-sulfur bituminous coal and lignite were demonstrated in 1980. Conventional pulverized fuel steam generators are more highly fuel-specific and typically lose capacity or availability when burning an off-design fuel.

- PFBC is amenable to bringing on-line relatively small incremental blocks of power with short lead times, and without sacrificing the capital-cost advantages of large central power stations. Though not yet fully quantified, there is some maximum single-combustor size, probably in the 150 MW_e range, that can be shop-assembled and transported via barge and/or rail to a large number of power plant sites. Benefits would include reduced construction time and commensurate reduced AFDC charges.

All of these benefits are available in varying degrees and with varying degrees of technical risk in the several commercially-proposed PFBC cycles. EPRI, in an effort to normalize the risk/benefit ratios of the cycles, contracted with Brown Boveri and Babcock & Wilcox in 1980 to conduct an engineering review of the cycles. The results of their work are described in the next sections.

PFBC POWER CYCLES

In order to take a fresh look at the practicability of PFBC, EPRI contracted with Brown Boveri and Babcock & Wilcox two years ago to analyze the thermal performance and engineering feasibility of the most commonly suggested PFBC cycles. BBC used consistent assumptions for the evaluations and normalized all of the cycles to their Model 11 gas turbine. The cycles are described below:

Cycle A: Steam Tubes In Bed, High Temperature Gas Turbine

This type of "conventional" cycle, generally associated with the PFBC/CC concept and favored by General Electric and Stal Laval, is depicted in Figure 3. The PFB steam tubes extract much of the heat from the bed and gas stream, yet the 1560°F exit temperature permits the gas turbine to generate about 50 MW_e, i.e., 20% of the total combined cycle output. As the turbine expansion drops the gas temperature to only 780°F, a heat recovery steam generator (HRSG) is required. The HRSG (1) heats the boiler feedwater supply to the PFB (economizer surface), and (2) provides superheated steam (which is mixed with the cold reheat steam supply prior to passing through the reheaters in the PFB).

This cycle has the highest gross thermal efficiency ($h_{th} = 41.4\%$) of all of the cycles studied, and, with a gas turbine inlet temperature of 1560°F, is comparable to a gasification plant with a turbine inlet temperature of 2300°F.

Cycle B: Turbo Expander

This case was conceptually predicated upon the use of an existing-technology, erosion-resistant, single-stage turboexpander. It would ideally be a machine similar to expanders designed to operate on fluid cat-cracker (FCC) regenerator off-gas. Such expanders, having gone through a three-decade design evolution period, are presently able to operate with entrained catalyst loadings of 120-160 ppm, inlet pressures of 2 to 3 atm., temperatures of 1200°F, and gas mass flows of 330 lb/sec. They have successfully achieved turbine blade lifetimes in excess of 25,000 hours. This operating period has been dictated by the economics of refinery operation, not blade wear, because the cost of a forced refinery outage due to an expander failure would far exceed the cost of routine conservative, preventative maintenance. Because such a turbine does not exist in the BBC product line, the project team chose to evaluate the use of a conceptual gas turbine in this cycle. The turbine would have a pressure ratio (PR) of 5:1, gas mass flows commensurate with the BBC Model 11 gas turbine and a combined turbine-compressor-mechanical efficiency equivalent to that of the BBC Model 11. Two variants of the cycle were analyzed: (1) a combined-cycle configuration (identical to cycle A, Figure 3) with a turbine inlet temperature of 1300°F (HT case), and (2) a self-sustaining case with a turbine inlet temperature of 615°F (LT case, identical to cycle E, Figure 6).

The thermal efficiency of the LT case and cycle E are the same while that of the HT case in less than one percentage point higher. The PR of B is 5:1, whereas the

PR of E is 10:1; hence, the size of the hot gas-side components in B will be about 65% larger than for cycle E (B operates at 615°F turbine inlet temperature vs. 770°F for cycle E). Note that the extremely conservative LT cycle B has a thermal efficiency about only two points less than the most efficient cycle A.

It should be noted that the efficiency of the gas turbine in these cases was specified to be 87.8%. The efficiency of existing cat-cracker turboexpanders, however, will generally be 80% or less because such machines are single-stage at PR of 2-3 and aerodynamic efficiency has been traded-off for erosion resistance. The thermal efficiency of either the combined-cycle or the self-sustaining turbocharged boiler will be reduced through the use of the less-efficient turboexpander.

Cycle C: Hot Air Tubes In Bed

This cycle, espoused by Curtiss-Wright, unlike Schemes A, B or E, is a "pure" combined cycle; i.e., only the exhaust of the coal-fired gas turbine is used to generate steam (Figure 4). As a result, the steam state conditions are low compared with cycles wherein steam is generated within the PFBC, and consequently the thermal efficiency suffers. The compressor air discharge is divided into two streams so that two-thirds of the total compressor air flow passes through the tubes of an in-bed heat exchanger and is raised to within 41F° of the bed exhaust gas temperature. The remaining 1/3 of the air is, of course, used for combustion. The heated, clean air is blended back into the exhaust gas from the PFB and the mixture is delivered to the gas turbine. The turbine exhaust generates steam in a heat recovery steam generator (HRSG). As in all "pure" combined cycles, the cycle thermal efficiency is a strong positive function of the turbine inlet and outlet temperature, and a moderate function of the turbine inlet and outlet temperature and pressure ratio. The optimum pressure ratio occurs at 8.5 atm.; however, the cycle was analyzed at 7 atm. to be comparable to the Curtiss-Wright cycle. At 7 atm. and 1600°F, BBC has calculated a thermal efficiency of 33.6%; although even at the optimum pressure ratio, the efficiency increases only to 34%.

These efficiency values are considerably lower than calculated by Curtiss-Wright (39.0%). Differences in technical assumptions account for the variations:

- C-W designed this system with a bilevel steam system with higher steam conditions than selected by BBC.
- C-W used less-conservative estimates of pressure loss through the air tubes and other air-side components than BBC.
- C-W used considerably less cooling air through the turbine than BBC.
- C-W used higher turbomachinery efficiencies.
- C-W used a lower condensing temperature.

The positive and negative implications of using air as the cooling medium for PFBC in-bed tubes are summarized below:

- Tube metal temperatures of f1600°F (vs. f1200°F for steam superheater tubes) present severe materials problems.
- Low heat transfer rates to air, as compared to steam, dictate the use of greater amounts of tube surface area. Complex and costly internally-finned tubes are under development.
- Plant efficiency drops off rapidly at partial loads because h_{th} is a strong function of turbine gas inlet temperature. This may be substantially obviated by the use of multiple PFBC modules, but at higher cost and complexity.
- Performance demands on the HGCU system are somewhat reduced from the steam cycle case because of the dilution of the gas stream by the clean hot air. The overall reduction in performance requirements via the dilution will not have a substantive impact on lessening the total gas cleanup demands.

- The air tube cycle would have about 1/3 the water consumption of the PFBC steam cycles because of reduced heat rejection quantities from the steam turbine condenser.

- As only 1/3 of the air is used for combustion the output per PFBC is relatively small, requiring more gas turbine/PBFC modules for a given total plant output.

- The cycle is very sensitive to gas turbine fouling and efficiency drop because such a large fraction of total plant output is from the gas turbines.

Cycle D: Adiabatic Cycle

There is no indirect cooling of the bed or combustion gases via tube banks in the adiabatic cycle case (Figure 5); the cooling duty in the bed is performed entirely with large volumes of excess air. This case also is a "pure" combined-cycle and, as in the previous case, the thermal efficiency suffers due to poor steam conditions. The level of particulate loading in the PFBC off-gas stream will be inherently lower than Schemes A, B or D (~6,000 ppm) because of the low rate of coal input per unit of air input. This low loading (~1/3 of the steam-cooled bed case) will not significantly reduce the performance demands upon the hot gas cleanup system.

Cycle E: Turbocharged Boiler; Low Temperature, Self-Sustaining Cycle

Cycle E (Figure 6) is a limiting case and variant of cycle A, wherein the turbine operates at a self-sustaining condition (zero net power output), analogous to diesel or gasoline engine turbocharging. All the power is developed by the steam turbine. As the gas has been cooled to the point where the turbine is just self-sustaining, the severity of the gas turbine operating conditions have been markedly decreased.

This low-temperature cycle, similar to low-temperature cycle B, but operating at 10 atm. and 770°F instead of 5 atm. pressure and 615°F, has a gross thermal efficiency of 39%, 2 percentage points less than the most efficient cycle A. BBC's analysis predicts that this cycle (E), albeit with a lower efficiency will at least achieve busbar cost of electricity parity with cycle A because of substantially lower capital costs and markedly higher first-generation reliability.

The lower capital cost is a result of a number of factors, including elimination of the heat recovery steam generator and the gas turbine generator (the heat is removed from the gas, and transferred to the steam, by additional convective heat transfer surface in the PFB). Because the gas temperature is below 800°F, substantial costs can be saved by using low-alloy components in the entire hot gas path and externally-insulated gas pipes. The mechanical design of the entire hot gas path is greatly simplified and, most importantly, near-conventional technology hot gas filters can be applied with confidence. With the application of a near-100% efficient filter, gas turbine fouling, corrosion and erosion are substantially eliminated and the baghouse on the gas turbine exhaust, needed to satisfy environmental regulations in the other cycles, is eliminated.

Lower capital costs in this plant are also a consequence of the increased gas density at the lower temperature (the hot gas in Cycle E is 1.7 times more dense than the gas in Cycle A). For constant velocity in the hot gas system, 40% fewer filter elements are required and the hot gas pipe diameter is decreased by a fourth, compared to the high temperature PFBC cycles at 10 atm. pressure. Substantial reduction in hot gas pipe wall thickness will result from the lower gas temperature and reduced pipe diameter. Figure 7 summarizes the key characteristics of Cycle A, C, D and E with respect to hot gas cleaning at both high temperature, high pressure (HTHP) and at the stack gas conditions. The "Gas Volume Treatment Requirement" columns, which indicate the quantity of gas to be cleaned to generate a kilowatt-hour of electricity, demonstrates the comparatively small amount of gas cleaning required for the turbocharged boiler cycle (last row) at HTHP conditions.

The turbocharged boiler has other advantages: (1) Control requirements are less stringent. Because there is no generator on the gas turbine, there is virtually no danger of loss of load and gas turbine overspeed protection is virtually eliminated. There is also no longer need for precise turbine speed regulation as the gas turbine can float freely during operation to satisfy boiler airflow demand. (2) The strength of materials at 770°F is substantially greater that at 1550°F. As a result, the mechanical design of both the hot-gas cleanup and hot-gas ducting systems can be considerably simplified. The key barrier that has prevented the successful development of a high pressure, high temperature (1550°F) filter has not been so much the performance of the filter media itself (several acceptable media are available), but the design of the structure to support the filter media. Because the gas density is so low at 1550°F, the filter unit is necessarily large and the design of the mechanical system becomes complex. In the PFBC turbocharged boiler cycle, low alloy, uncooled steel members can be used in considerable simplified mechanical designs of both systems with high confidence. (3) Gas turbine reliability uncertainty is substantially reduced; thus, power loss due to erosion and fouling is reduced to near zero due to the application of the filter, corrosion problems are avoided because all alkali eutectics are solid phase at 770°F and existing turbo-expander technology can be applied with minimal modification.

It is this cycle, the PFBC turbocharged boiler, that EPRI has decided to pursue, chiefly because of attractive economics, the high conservatism/low risk of the cycle and the existence of two near-term applications (described later).

EXPERIENCE WITH TURBOCHARGED BOILERS

The concept of a combustor and boiler operating at high pressure is not new and that concept has extensive operating experience--at least with heavy fuel oil.

The most extensive experience record belongs to Brown Boveri. BBC, from the early 1930's through 1976, built 112 "Velox" boiler plants ranging in size up to 35 MW_e that accumulated in excess of 5 million hours of operating time. All BBC plants were built for midrange- and peak-load operation and were designed for low capital cost. Because the BBC Velox boilers run with a turbocharger inlet temperature of about 800°F, they have operated very successfully while burning heavy fuel oil.[8]

Foster Wheeler built 35 turbocharged boilers for use as Naval ship power plants in the late 1950s and the 1960s. The important benefits realized by the Navy were 50-percent reduction in both weight and volume of the boiler as compared to conventional units. After curing some structural problems, the first units achieved the predicted thermodynamic performance.

PFBC R&D GOALS AND UTILITY NEEDS

When worldwide development of PFBC began in the late 1960s, the utility industry was focusing its technical efforts on maintaining the historic trend of declining average unit production cost. That decline had been achieved by the introduction of improved technologies, the capture of increasing scale economies, and the retirement of older, less efficient units. PFBC development programs, established to capitalize upon the benefits listed in the preceeding sections, were structured to fit industry goals of the day and emphasis was placed on developing highly efficient, 600-1000 MW_e PFBC plants.

The operating environment of the electric utility industry drastically changed in the 1970s. New factors which did not exist before have significantly affected utility operating and expansion strategies. These include:

- oil price shocks,
- escalating cost of money,
- unpredictable and regionally-variable demand growth that is substantially below historic levels,
- increasingly active environmental special-interest groups,
- increasing regulatory intervention;
    - economic penalties for over-capacity,
    - increasing plant lead-times,
    - substantial environmental compliance costs,
    - declining viability of the nuclear option,
- failure of utility commissions to grant adequate rate relief.

As a consequence of these factors, utilities are making smaller capital investments and have halted or decreased capacity expansion construction programs (over half of the new plants on order that were scheduled to come on line by 1989 were either cancelled or delayed in the April 1980-April 1981 time period). The effects can be observed in the trend toward smaller new coal-fired unit size, the growing utility interest in up-rating the capacity of existing units and the declining capacity factor for gas- and/or oil-fired units.

PFBC researchers to date have generally not factored the recent drastic changes in the utility operating environment into their development objectives. It now appears that two comparatively new market niches exist as a consequence of current utility business conditions that could be filled by PFBC turbocharged boilers. These market niches are described in the next two sections.

FOSSIL PLANT UPRATING

The industry is seeking to bring on line the lowest incremental cost capacity it has available. This currently appears to be the capacity available by uprating existing steam turbine/generators. Studies by Westinghouse and others[9,10,11,12] have shown that the turbine/generators in over half of the nation's 1250 coal-fired units are capable of being uprated an average of 11%. The total latent turbine/generator capacity available by uprating is estimated to be 15 GW_e, and is achievable through the application of materials, technology, and design improvements that were not available when the units were originally built. Changes to the turbine/generator would most frequently include nozzle block modifications and generator rotor and stator rewinding. Components required for the modification would be available within two years and could be installed during regularly scheduled shut downs.

Many of the units available for uprating were manufactured in the 1960s and early 1970s, a highly competitive period for the nation's boiler manufacturers. The boilers installed then were carefully engineered to reduce costs and, as result, surplus performance margins were cut to zero. As unit design was cut closer and closer, units were frequently unable to achieve full rating. This problem has been exacerbated through the intervening years because of declining coal quality. Hence, the boiler will likely become the limiting variable in the effort to capture the latent capacity in many existing turbine/generators. Westinghouse has recognized this problem and together with the nation's boiler manufacturers, has investigated the feasibility of adding comparatively small boilers to the plants that cannot uprate their existing boilers. Current interpretations of the

New Source Performance Standards indicate that the supplemental boilers would require SO_2 scrubbers. This requirement, especially at a non-scrubber site, would dampen utility acceptance of the concept.

The cost of incremental capacity available from uprating, even if a supplemental boiler is required, is estimated by Westinghouse to be about half of that required for new plants.

The PFBC turbocharged boiler could suit this application very well because of rapid delivery and erection of shop assembled steam generator components, the low areal requirements of the turbocharged boiler and the elimination of the scrubber requirement.

REPOWERING

The repowering of either low capacity factor, premium-fuel burning fossil plants or retired fossil plants with new PFBC turbocharged boilers is also a potentially large market for first-generation plants. American Electric Power, Boston Edison Company, Public Service Electric and Gas, and El Paso Electric Company have all conducted preliminary investigations of PFBC combined cycle repowering of existing stations. As the cost of premium fuels rises and these units are displaced from service, they are expected to become more favorable candidates for repowering with coal-fired boilers located adjacent to the existing gas- and/or oil-fired unit.

COMPONENT TRANSPORTATION

Meeting the requirements of these two market niches largely depends on being able to transport the major plant components from the shop to the field. Recent work by EPRI has investigated the traffic issues of moving the components on the nation's inland waterway system. To establish a conceivable upper limit for transportation feasibility, the study was based upon the movement of an existing Foster-Wheeler PFBC design (115 ft. overall length, 36 ft. diameter, 1300 ton net weight, and 125 MW_e capacity). This design was originally done during the General Electric Coal-Fired Combined-Cycle DOE program, and was conceived to be field erected using standard power industry and petrochemical refinery industry methods. It was rationalized that if this design could conceivably be transported, then shop assembly and transport would indeed be feasible. It was found that this PFBC configuration could indeed be moved over 16,000 miles of the nation's inland waterways and could reach 122 existing coal-fired stations totaling 100 GW_e of capacity.

THE TOTAL PACKAGE

Combining the existing unit uprating concept and/or the repowering concept with the shop-assembled and transported PFBC turbocharged boiler offers one integrated solution to the utility problems discussed earlier. The same basic package of benefits available from the uprating or repowering of plants with conventional pulverized-coal-fired boilers (lowest capital cost for the incremental capacity, premium fuel backout, extremely rapid lead time, favorable cash flow, and ability to precisely match load growth) would, of course, be available. However, extra benefits would be available with the use of PFBC turbocharged boilers (instead of PC/FGD boilers) because of the ability to manufacture these units in the shop and transport them to the site. Substantial reduction in field erection time, and the attendant reduced finance charges and improved cash flow effects, are expected to be achieveable with the PFBC turbocharged boiler. SO_2 scrubbers would also not be required on the PFBC turbocharged boiler, thereby mitigating resistance to the idea of adding new boilers to a scrubber mandated plant.

From a developmental needs point of view, the uprating concept is particularly attractive because the PFBC turbocharged boiler unit could be built without stringent requirements for turndown capability. The parent unit of the duo could accomodate load swings, leaving the PFBC turbocharged boiler to operate steady state at full load, thus reducing developmental risk even further.

EVOLUTION TO COMBINED CYCLES

The concept of uprating the capacity of conventional plants with PFBC turbocharged boilers is attractive from both the utility and PFBC development viewpoints. From the utility perspective, extra capacity can be available at already permitted sites, at low cost and with short lead-times; from the PFBC development perspective, a utility need exists that can be met with the low-risk, PFBC turbocharged boiler. In addition, it is an undemanding application with regard to turndown and rate of load change making it ideal for first-generation plants.

These first-generation plants will conclusively resolve the fundamental uncertainty that exists today regarding the ability to operate gas turbines on cleaned coal combustion products. It is this resolution, demonstrated in this low-risk configuration, that will induce designers, vendors, owners, and operators of future PFBC plants to begin increasing the turbine "firing" temperature (thereby taking net power from the gas turbine) to combined-cycle conditions.

PFBC R&D PLAN: 1982-1987 *

The previous sections described the main benefits available from PFBC, the chief PFBC cycles of interest, the rationale for PFBC development in the 1970s and why the unique circumstances in the 1980s have caused the technical staff to emphasize the PFBC turbocharged boiler as an opportunity to accelerate the development and commercial application of this technology. This section describes the work elements of the redirected PFBC subprogram that are required to position EPRI to begin a PFBC turbocharged boiler prototype plant project.

The subprogram is comprised of five major work categories (Figure 8). Category 1, "Existing R&D Programs," contains all of the work presently under contract that was started when the interest of the subprogram was on high temperature combined-cycles. The majority of this work responds to the needs of both high- and low-temperature PFBC cycles and will be continued to its currently planned termination.

To accurately define the potential cost savings of shop assembly and transport of turbocharged boilers, "Transportability and Shop Fabrication Studies" (Category 2), will be conducted early in the revised program. These will define the approximate cost and schedule savings associated with shop fabrication of modules, determine the upper limits on the size of the modules, define the number of power-plant sites that can be reached by barge, and define the technology required to unload barges at powerplant sites.

In the "Commercial Plant Design and Economics Studies" (Category 3), teams of contractors will prepare commercial designs of the baseline configuration with modular construction and estimate the total costs of generating electric power

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* Personal Communication with S.G. Drenker, August 4, 1983, reveals that due to budgetary reductions the timing of the events described in this section are subject to delay and uncertainty. A revised PFBC R&D plan is being prepared by EPRI.

compared to a PC/FGD plant. PFBC technology will include conventional bubbling bed combustors as well as circulating PFBC's. All of the cost and design studies will be normalized by an architect/engineer to assure comparability. Also, all new designs and design modifications will be compared with the baseline design using EPRI economic models. The commercial plant studies will start in early 1983 and continue for approximately 30 months. Results of these studies will determine whether PFBC technology can be commercialized in a shop-assembled, barge-transported configuration.

After the commercial studies are underway, a new series of "Technology Support Projects" (Category 4) will be pursued which directly support commercialization of the baseline technology. Some of these programs will be extensions of existing technology projects. All of the data will be incorporated into the commercial studies to assure realistic results. Category 5, "Prototype Plant Project; Preparatory Activity," includes preliminary feasibility analyses of various projects.

The prototype plant (Categories 5 and 6) will preferably be a repowering application of an existing plant. Solicitation of utility host sites and cosponsors for the prototype plant will begin in 1984, based on the results of the preceeding development. Studies will be conducted concurrent with the commercial plant studies, increasing in effort as input is obtained from the commercial plant studies and as utility interest is heightened by economic results. Site studies will be completed in late-1984 when the commercial plant studies are essentially complete and the technology support program has provided substantial confirmation of the technology's feasibility. After six months of evaluation by EPRI, a final selection will be made.

The PFBC development plan described above offers substantially reduced risk compared with previous plans. This reduced risk also implies reduced cost to develop, construct, and operate the first generation plants and a lengthened, although now realistic, schedule to achieve the ultimate goal of combined-cycle plants. The R&D plan proposed here also is predicated upon the successful operation of the as-built and as-funded Grimethorpe rig. If the IEA plant fails to produce the operational data it is designed to produce, then the EPRI schedule will be delayed accordingly. Other PFBC pilot facilities are also now completing construction. These are the Stal Laval facility in Sweden and the Curtiss-Wright large pilot plant. These should also add to the confidence necessary to implement the proposed turbocharged boiler prototype.

CONCLUSION

The operating environment of the electric utility industry in the United States changed substantially in the 1970s. Declining industry financial health, uncertain demand growth and increasing regulatory intervention have forced the industry to reduce its generation expansion programs. More and more utilities are emphasizing the use of smaller coal-fired power plants, extending the life of existing coal-fired power plants, and increasing the capacity of existing coal-fired power plants.

Past PFBC R&D concentrated on highly efficient, large baseload combined-cycle power plants. The technical barriers to this technology, primarily hot-gas cleaning, hot-gas ducting and gas turbine erosion, fouling, and corrosion, are being brought down, but confidence in commercially-applied PFBC plants is still limited. By shifting the course of PFBC development toward the turbocharged boiler, the technology can be commercialized on an accelerated schedule with reduced risk and can also be economically applied to utility plant uprating or repowering in the late 1980s and early 1990s. Uprating of existing power plants to raise total plant output by adding supplemental PFBC turbocharged boilers promises to be the lowest cost incremental capacity available in that period. By developing PFBC in this low-risk configuration and proving its feasibility in financially attractive repowering applications, sufficient confidence will be gained to increase the firing temperature in future plants to combined-cycle conditions. This evolutionary path will eventually lead to the 40%+ efficient, direct coal-fired, combined-cycle power plant.

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Power Plant*	Heat Rate Btu/kWhr
PFBC Combined Cycle	8,470
PC/FGD (wet limestone)	9,860
PC/FGD (regenerative	10,290
AFBC	9,640
Gasification (Texaco) combined cycle	9,460

* 2400 PSIA, 1000°F, 1000°F current technology.
Source: EPRI Technical Assessment Guide, May, 1982

Figure 1. Coal fired power plant heat rate comparison.

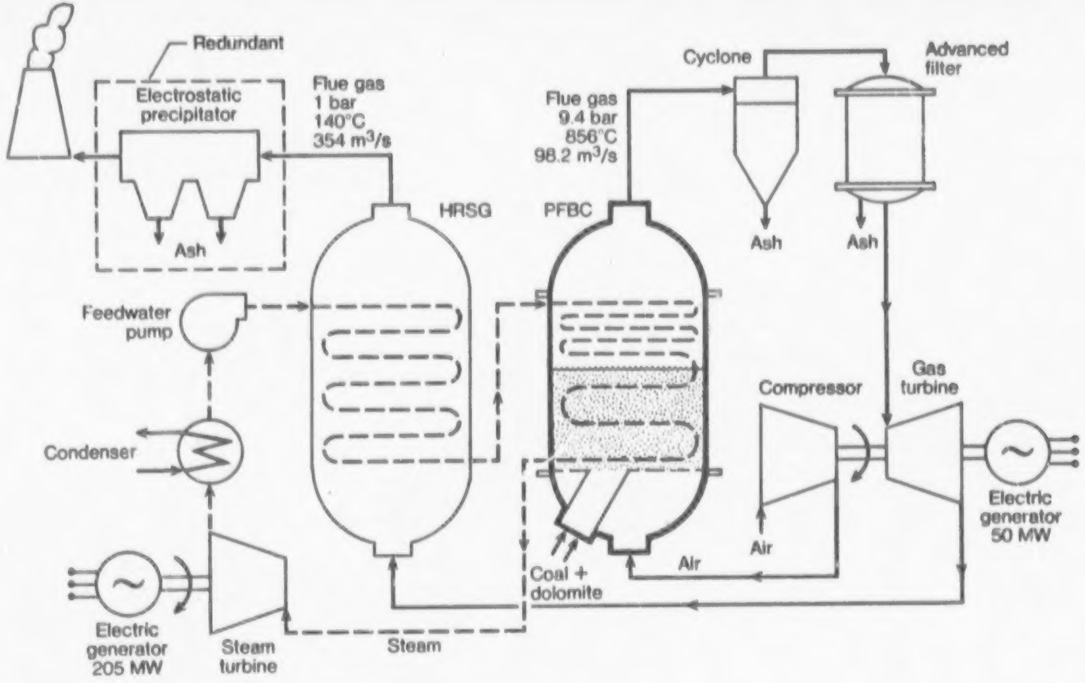

Redundant
Electrostatic precipitator
Ash
Flue gas
1 bar
140°C
354 m³/s
HRSG
Feedwater pump
Condenser
Electric generator 205 MW
Steam turbine
Steam
Flue gas
9.4 bar
856°C
98.2 m³/s
PFBC
Coal + dolomite
Air
Cyclone
Ash
Advanced filter
Ash
Compressor
Air
Gas turbine
Electric generator 50 MW

Figure 2. Simplified PFBC Combined-Cycle Diagram

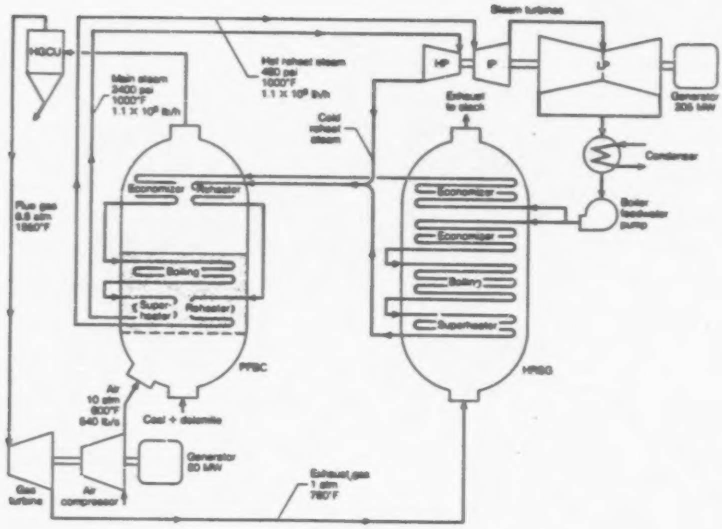

Steam turbines
HGCU
Hot reheat steam
460 psi
1000°F
1.1 × 10⁶ lb/h
Main steam
2400 psi
1000°F
1.1 × 10⁶ lb/h
HP
IP
LP
Generator
205 MW
Exhaust to stack
Cold reheat steam
Condenser
Economizer
Reheater
Boiler feedwater pump
Flue gas
6.8 atm
1560°F
Economizer
Boiling
Boiling
Super-heater
Reheater
Superheater
PFBC
HRSG
Air
10 atm
600°F
640 lb/s
Coal + dolomite
Generator
80 MW
Exhaust gas
1 atm
780°F
Gas turbine
Air compressor

Figure 3. High temperature PFBC combined-cycle. Cycle A.

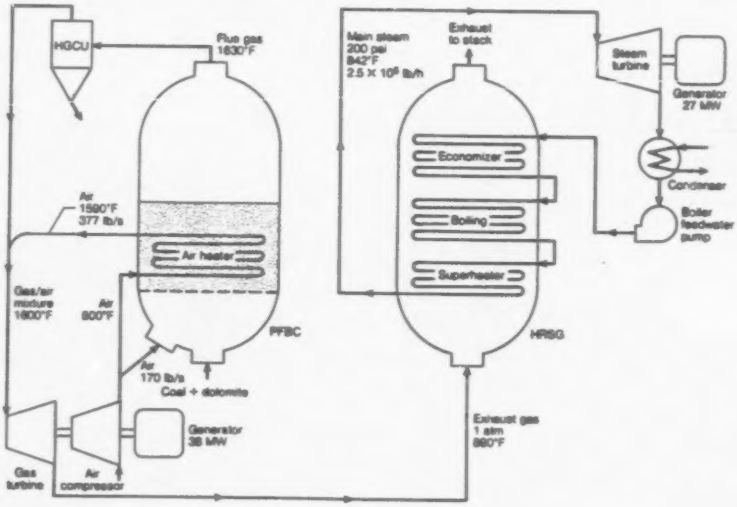

HGCU
Flue gas
1630°F
Main steam
200 psi
842°F
2.5 × 10⁵ lb/h
Exhaust to stack
Steam turbine
Generator
27 MW
Economizer
Condenser
Air
1590°F
377 lb/s
Boiling
Boiler feedwater pump
Air heater
Superheater
Gas/air mixture
1600°F
Air
600°F
PFBC
HRSG
Air
170 lb/s
Coal + dolomite
Generator
38 MW
Exhaust gas
1 atm
860°F
Gas turbine
Air compressor

Figure 4. Air heater PFBC combined-cycle. Cycle C.

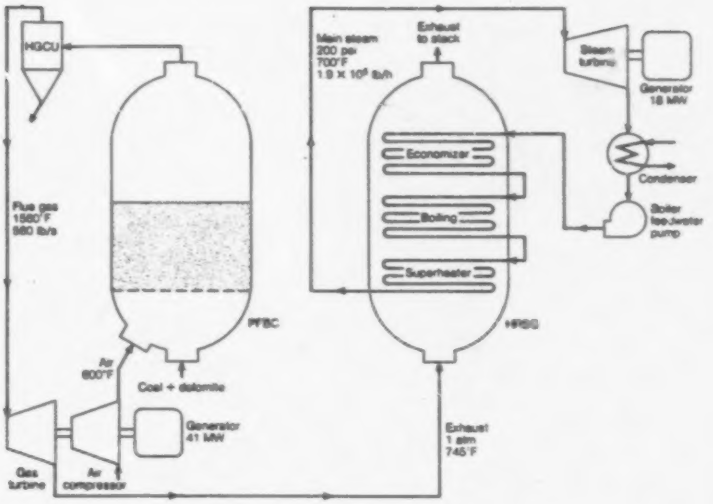

HGCU
Flue gas
1560°F
560 lb/s
PFBC
Air
600°F
Coal + dolomite
Generator
41 MW
Gas turbine
Air compressor
Main steam
200 psi
700°F
1.9 × 10⁵ lb/h
Exhaust to stack
Steam turbine
Generator
18 MW
Economizer
Condenser
Boiling
Boiler feedwater pump
Superheater
HRSG
Exhaust
1 atm
745°F

Figure 5. "Adiabatic" PFBC combined-cycle. Cycle D.

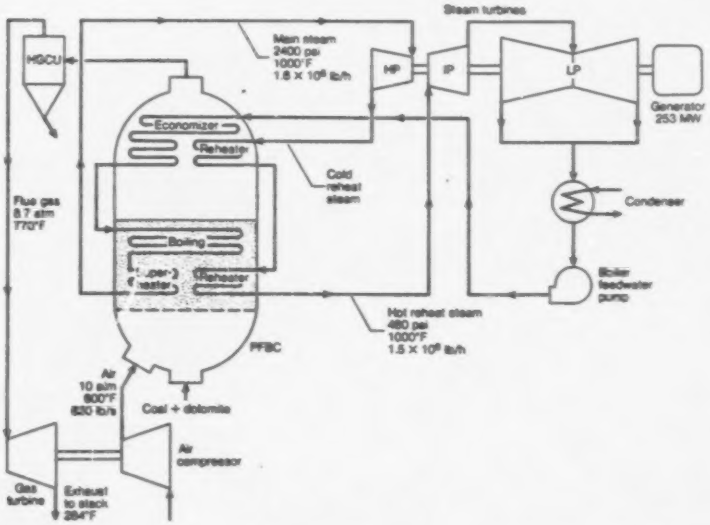

Steam turbines
HGCU
Main steam
2400 psi
1000°F
1.8 × 10⁶ lb/h
HP
IP
LP
Generator
253 MW
Economizer
Reheater
Cold reheat steam
Flue gas
8.7 atm
770°F
Condenser
Boiling
Super-heater
Reheater
Boiler feedwater pump
Hot reheat steam
480 psi
1000°F
1.5 × 10⁶ lb/h
PFBC
Air
10 atm
600°F
620 lb/s
Coal + dolomite
Air compressor
Gas turbine
Exhaust to stack
284°F

Figure 6. Turbocharged boiler PFBC simple cycle. Cycle E.

Cycle	Cycle Description	Thermal Efficiency (%)	Gas Turbine		Power Output					Specific Work (kW(e)/lb air/s)	Steam Conditions			Gas Conditions at HGCU			Gas Volume Treatment Requirements Based on Power Output From		Sensitivity of η_{th}* to G.T.	
						S.T.		G.T.			Main Steam		Reheat							
			Pressure Ratio	Inlet Temperature (°F)	Total (MW(e))	(MW(e))	(%)	(MW(e))	(%)		Pressure (psig)	Temperature (°F)	Temperature (°F)	Density (lb/acf)	Specific Volume (acf/lb)	Specific Volume Standardized Against Cycle A	Gas Turbine (acf/kW-h)	Total Plant (acf/kW-h)	Inlet Temperature	Pressure Ratio
A	Steam tube in bed, high temp. G.T.	40.0	10:1	1580	286	228	80	58	20	92	2400	1000	1000	0.18	5.5	1.0	239	46.6	Low	Low
B-1	Turboexpander, high temp.	38.6	8:1	1300	252	219	87	33	13	58	2400	1000	1000	0.10	9.8	1.8	684	66.5	Low	Low
B-2	Turboexpander, low temp., self sustaining	38.0	8:1	800	253	253	100	0	0	0	2400	1000	1000	0.17	6.0	1.1	—	83.4	Independent	Low
C	Air tube with hot air/gas blending and HRSG	33.6	7:1	1600	65	27	42	38	58	70	200	842	—	0.12	8.7	1.6	140	87.0	High	Medium
D	Adiabatic with HRSG	34.5	10:1	1580	59	18	31	41	69	73	200	700	—	0.18	5.5	1.0	268	186.0	High	Medium
E	Steam tube in bed, low temp., self sustaining	38.0	10:1	770	253	253	100	0	0	0	2400	1000	1000	0.30	3.4	0.6	—	30.0	Independent	Independent

*η_{th} denotes cycle thermal efficiency

Figure 7. Comparison of major PFBC thermodynamic cycles.

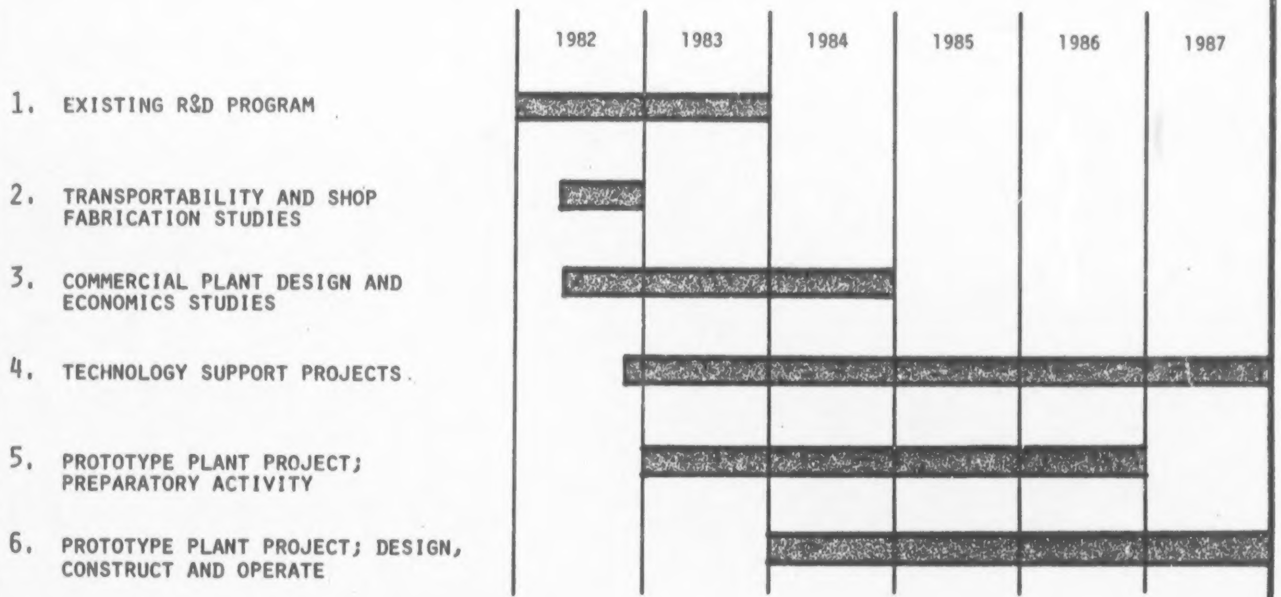


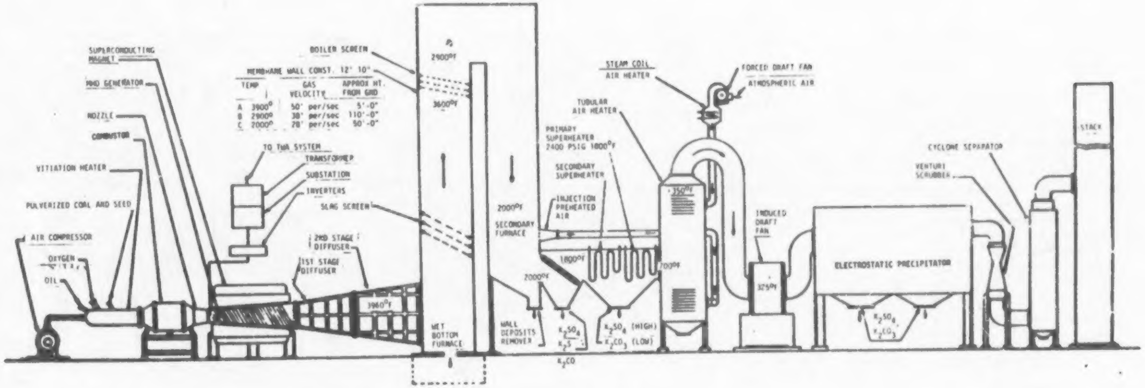
Figure 8: PFBC SUBPROGRAM PLAN SUMMARY

A more efficient way to burn fossil fuels, magnetohydrodynamics (MHD) produces electricity directly from thermal energy, by-passing the conversion to mechanical energy. MHD burning efficiency of fossil fuels is expected to be about 50%, as compared to only 33-40% for convention power generation systems. This section is taken from a 1979 Mitre report.

OPEN-CYCLE MAGNETOHYDRODYNAMICS

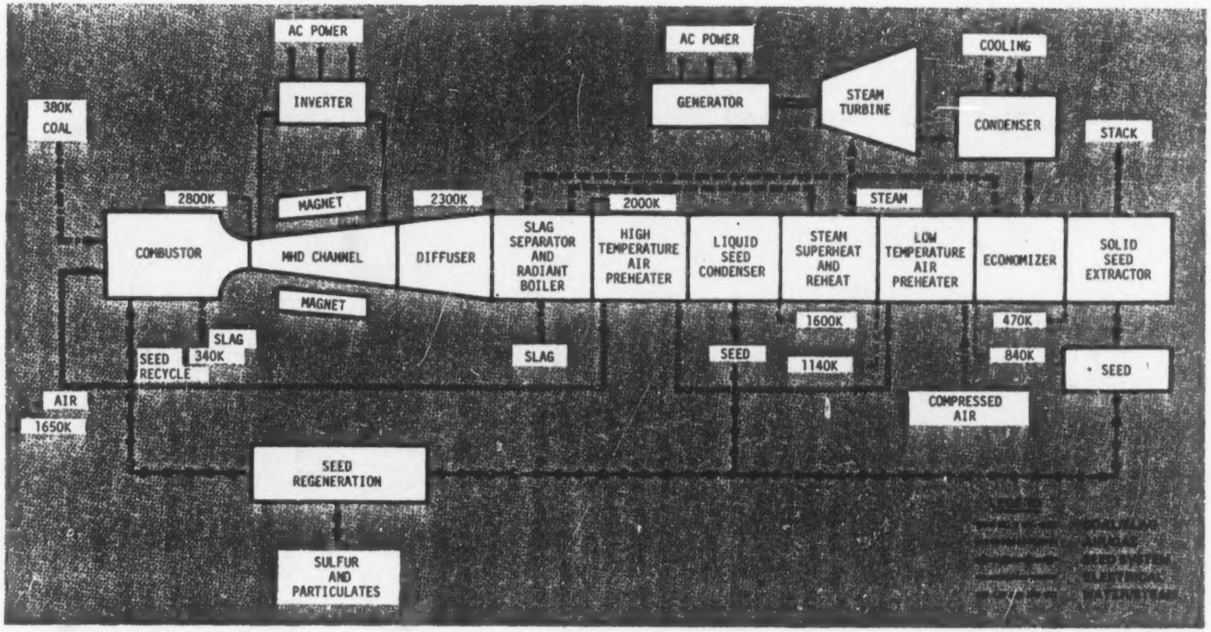
The Open-Cycle Magnetohydrodynamics (OCMHD) converts heat energy of coal to electricity directly without the need for flue gas desulfurization, and its successful development can lead to substantial increases in overall plant efficiency. However, to obtain the high overall plant efficiency, OCMHD must be used as a topping cycle to another heat conversion system because of the relatively low enthalpy extraction of MHD generators. At present, conventional steam bottoming is proposed.

The MHD power plant consists of two distinct subsystems, the MHD topping section and the steam bottoming section. As shown in Figure 2, pulverized coal is burned in the combustor with a fuel-to-air ratio of 1:1. The fuel-rich combustion minimizes the probability of nitrogen oxides formation. However, to obtain the high operating temperature (4800°F), for proper operation, the combustion air must be preheated to about 2500°F. Alternately, air enriched to various oxygen levels could be used with modest levels of preheating, up to 1100°F, using any coal. A large portion of the slag (85 percent) is tapped out of the combustor. The remaining portion is transmitted


SUPERCONDUCTING MAGNET
MHD GENERATOR
NOZZLE
COMBUSTOR
VITIATION HEATER
PULVERIZED COAL AND SEED
AIR COMPRESSOR
OXYGEN
OIL
BOILER SCREEN
MEMBRANE WALL CONST. 12' 10"
TEMP
GAS VELOCITY
APPROX HT. FROM GRD
A 3900° 50' per/sec 5'-0"
B 2900° 30' per/sec 110'-0"
C 2000° 20' per/sec 50'-0"
TO TWA SYSTEM
TRANSFORMER
SUBSTATION
INVERTERS
SLAG SCREEN
2ND STAGE DIFFUSER
1ST STAGE DIFFUSER
3960°F
2900°F
3600°F
WET BOTTOM FURNACE
2000°F
SECONDARY FURNACE
WALL DEPOSITS REMOVER
2000°F
K_2SO_4
K_2S
K_2CO
1800°F
INJECTION PREHEATED AIR
SECONDARY SUPERHEATER
PRIMARY SUPERHEATER 2400 PSIG 1800°F
K_2SO_4 (HIGH)
K_2CO_3 (LOW)
TUBULAR AIR HEATER
350°F
700°F
STEAM COIL
AIR HEATER
FORCED DRAFT FAN
ATMOSPHERIC AIR
INDUCED DRAFT FAN
325°F
ELECTROSTATIC PRECIPITATOR
K_2SO_4, K_2CO_3
VENTURI SCRUBBER
CYCLONE SEPARATOR
STACK

SOURCE: Dicke 1978.

FIGURE 1
CONCEPTUAL PLANT ARRANGEMENT


AC POWER
INVERTER
380K
COAL
2800K
MAGNET
COMBUSTOR
MHD CHANNEL
MAGNET
SLAG
340K
SEED
RECYCLE
AIR
1650K
2300K
DIFFUSER
SLAG
SEPARATOR
AND
RADIANT
BOILER
SLAG
2000K
HIGH
TEMPERATURE
AIR
PREHEATER
AC POWER
GENERATOR
STEAM
TURBINE
LIQUID
SEED
CONDENSER
SEED
STEAM
SUPERHEAT
AND
REHEAT
1600K
1140K
STEAM
LOW
TEMPERATURE
AIR
PREHEATER
COMPRESSED
AIR
COOLING
CONDENSER
ECONOMIZER
470K
840K
STACK
SOLID
SEED
EXTRACTOR
SEED
SEED
REGENERATION
SULFUR
AND
PARTICULATES

SOURCE: Fossil Energy Research Program of the Energy Research and Development Administration, FY 1978, ERDA 77-33, 1977.

FIGURE 2
OPEN-CYCLE COAL-FIRED MHD SYSTEM

through the system with the combustion products. Although the temperature at the exit of the combustor and entrance to the MHD section is relatively high (>4800°F), the gas is still not sufficiently ionized. A seed must be introduced at this point to enhance its conductivity. The seed proposed is potassium carbonate (K_2CO_3). Approximately 1 percent, by weight, of seed in the gas is required to achieve the proper plasma conductivity.

The seeded gas and remaining slag enter the MHD channel, where the gas is expanded from approximately 8-to-10 atmospheres at the inlet to little more than 1 atmosphere at the outlet. The MHD channel is rectangular in cross section. A magnetic field from a magnet (superconducting) enters and exits from two opposite sides, and the other two sides are equipped with electrodes to collect the current generated from the interaction of the moving conducting gas and the magnetic field. This d.c. current is inverted to a.c. and fed into the power grid. The seed performs another interesting function: it combines with the sulfur in the coal to produce potassium sulfate, thus preventing the formation of oxides of sulfur (SO_x).

The combustion gases that are still at high temperature (4000°F) then enter the air preheater where additional slag is removed. It should be noted, however, that there is still uncertainty on how to build such an air preheater. Also, early commercial plants might not have a directly fired high-temperature air preheater. Following the air preheater is a standard steam generating plant (3600 psi/1000°F/1000°F). Instead of using a steam generating plant, the thinking in current research is to provide steam pressure of about 2400 psi in order to use the cooling water from the combustor, generator and diffuser. Additional air is introduced in a low-temperature air preheater to burn the remaining fuel. Following this is the steam economizer, a precipitator, and the stack. The seed is recovered as solids after the steam superheater and as a solid dust in the precipitator which precedes the stack. Because of the relatively high cost of the seed, it is recovered for reuse and for certain processes; pure sulfur is a product of this operation (Farah and Harlow 1976).

Process Chemistry

Prior to burning, coal is dried and crushed before being fed to the combustor. The water-cooled combustor operating with preheated air produces combustion gas products at temperatures in excess of 4600°F. The coal is seeded with potassium carbonate (K_2CO_3), alone or mixed with potasium sulfate (K_2SO_4), to achieve adequate levels of electrical conductivity in the combustion products. The mixture is burned in the combustion chamber with oxygen-enriched air. The K_2CO_3 added as a seed will react with SO_2 formed from the sulfur in the coal, eliminating the need for expensive sulfur-removal systems. Following recovery of K_2SO_4, a process is required to regenerate K_2CO_3 and produce a sulfur product that can be disposed of. There are several recovery processes, however none is established as best suited for MHD. Seven potential processes have been evaluated. Table 1 summarizes the advantages, disadvantages, and commercial status of these potential recovery techniques.

Technology Status

The principal difficulties of the OCMHD technology arise from the fact that the working fluid must have a sufficiently high conductivity to allow efficient operation. To obtain sufficient ionization in the working fluid, the temperature must be high; however, even stoichiometric temperatures are not sufficient to result in a high-conductivity plasma, and seed must be added. This creates the principal obstacle to commercialization: the high temperature requires materials that must withstand it. The fact that there are no moving parts in the generator portion of the MHD system relieves some of the requirements on the materials, but the fact that the system must operate in a utility grid for long periods with only regular maintenance intervals requires that the materials have long life under the erosive and corrosive action of the working fluid and seed/slag mixture. The presence of the seed/slag mixture in the steam plant further complicates the design of the superheater and boiler in this plant. In the MHD channel, where electrodes are imbedded in the walls and electrically isolated from each other by ceramic insulation, the requirement is for 2,000 to 5,000 hours of electrode life as well as for long-life insulators and seals between the two. Besides addressing the problems of materials and seed recovery, there is a need for efficient and economic methods for designing and constructing large high-field superconducting magnets.

To date, experimental MHD generators have been built and the most important accomplishments are:

- 100 hours of continuous operation at a relatively low 100 kW power range with moderate slag carryover and improved recovery of seed and slag from the gas stream
- 550 hours in channel test at V.S. (AVCO) with continuous periods of 250, 150, and 150 hours
- 250 hours in the MW power range with a clean fuel--this was accomplished on the USSR's U-25 generator (power placed in the Moscow grid)
- one-half hour without slag at 20.4 MWe on the U-25

TABLE 1

EVALUATION OF SEED RECOVERY PROCESSES

PROCESS	TECHNICAL EVALUATION		COMMERCIAL STATUS
	Pro	Con	
Engle-Precht	No thermal reduction	Lose one mole $MgSO_4$ for each mole K_2SO_4 High potassium loss ($\approx$40%). Potential water pollution problems.	Commercial process in Germany
Double Alkali	No thermal reduction	Dilute solutions. Large evaporative load, 50×10^6 lb H_2O/hr for 1000 MWe.	Commercial process for conventional power plants but modifications necessary for MHD
Formate	Simple process using common materials $Ca(OH)_2$	CO utilization unknown. May require coal gasifier. CO_2 must be removed.	Commercial process in Germany, 1930s
PERC	Dry conversion	Low CO and CO_2 utilization.	Pilot plant experiments
Aqueous Carbonate	Operations demonstrated	Non-commercial equipment. Design based on sodium not potassium.	New type equipment being developed by Atomics International, EPA, Empire State Electric Energy Research Corp.
Tomlinson-Tampella	Commercial process	Design based on sodium not potassium.	Commercial
Direct Reduction	One step process. Dry Operation. Material handling minimized.	Not developed.	Not developed

SOURCE: Matty et al. 1979.

- one minute at 32 MWe on the U-25, also without slag
- 400-hour test on materials for direct-fired preheater application with temperatures above 3000°F

In summary, the major accomplishments in the MHD program inlude:

- successful completion of $20MW_t$, 500-hour endurance tests under electrode loading conditions simulating commercial service and achieving order-of-magnitude improvement in electrode corrosion/erosion resistance
- successful subsonic generator operation at a high-magnetic field (5 tesla), representing the largest ($30\ MW_t$) high-field test ever run
- slag rejection in excess of 95 percent with carbon conversion of 97 percent achieved in a first-stage prototype of a two-stage combustor
- achievement of 95 percent sulfur removal from high-sulfur coal as well as NO_x reduction to 20 ppm
- 1,000 hours of cumulative testing at 2700°F on a direct-fired regenerative air preheater element under heat transfer, fluidynamic, and seed/slag conditions simulating MHD service
- design principles confirmed for large super-conducting magnets by operation of a 40-ton, 5-tesla magnet at design conditions

From an energy production standpoint, the enthalpy extraction of present experimental facilities has been somewhat less than predicted but within the acceptable range for the unit size. This has been attributed to the small scale of the experiments where wall losses are excessive. It is expected that at utility-plant scale (1,000 MW), the needed high enthalpy extraction will be achieved.

Major anticipated accomplishments for FY 1980 include:

- completion of facility activation at the Component Development and Integration Facility in Montana (CDIF)
- testing operation at the Coal-Fired Flow Facility (CFFF) at the University of Tennessee Space Institute

- channel testing at the Soviet U-25 facility with the U.S. channel to be delivered in late FY 1979
- testing in the High Performance Demonstration Experiment (HPDE) at the Arnold Engineering Development Center (AEDC) (U.S. Department of Energy 1979)

Commercialization Barriers

In addition to the technical barriers mentioned above, there are some major nontechnical barriers that must be overcome before MHD becomes commercialized. These arise because MHD plants are a radically new technology and that they may need to be in 1,000 MWe range for economical, efficient operation.

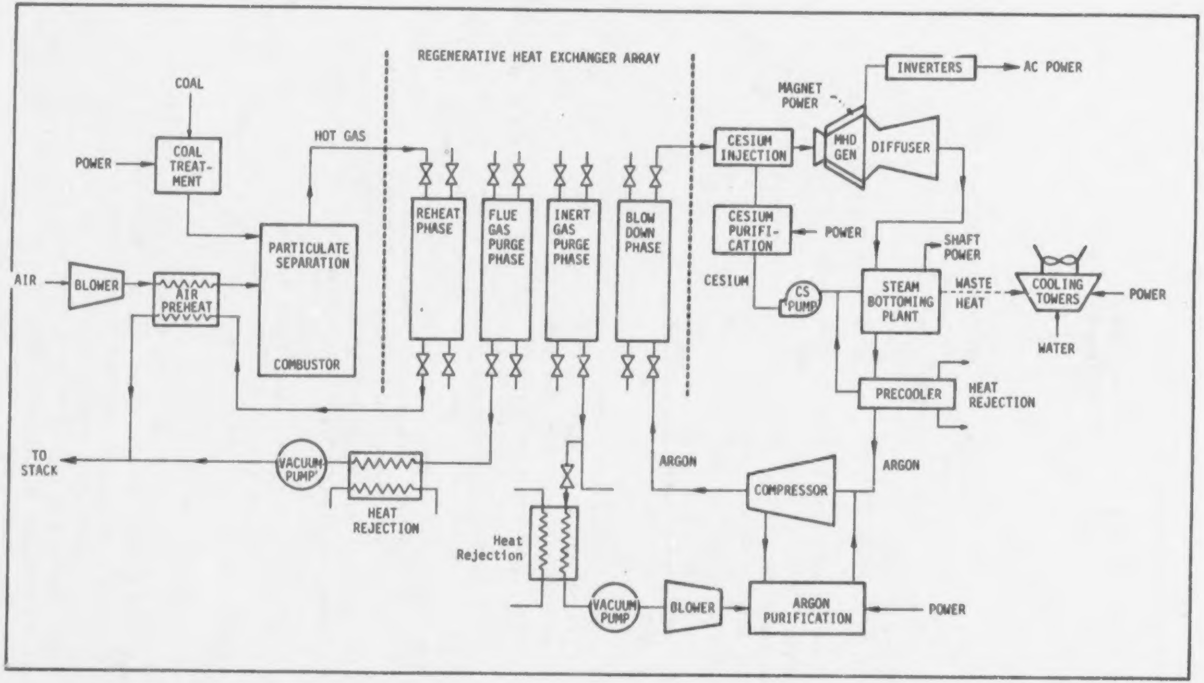
To citizen groups, large plants mean large land areas, hence large environmental intrusion. The MHD plant will require water, a transportation system for the coal, electrical substations, etc. In addition, the utilities will be asked to depend on the operation of new concept plants for a large portion of their power.

CLOSED-CYCLE MHD

The closed-cycle MHD processes being investigated are the closed-cycle plasma and the closed-cycle liquid metal. The basic energy conversion process of these processes is the same as that of the open cycle, namely, motional electromagnetic induction. However, the closed-cycle processes use a working fluid in a closed-loop system and receive the heat energy indirectly from a primary source through a heat exchanger. The primary heat source can be obtained from combustion of coal or other fossil fuels, or from a nuclear reactor if a suitable one is available. Because of closed cycling of working fluid, there is more latitude available in selecting a working fluid and in obtaining electron densities to give sufficient conductivity, resulting in lower required temperatures than for open-cycle MHD.

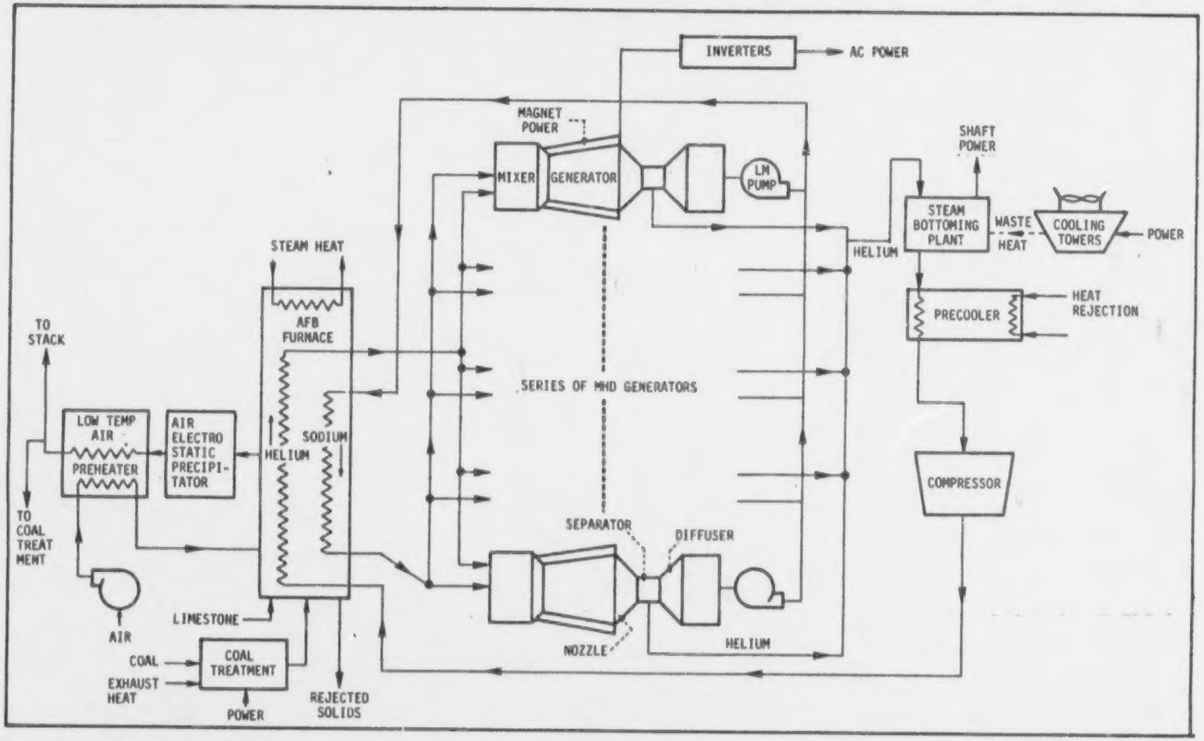
The closed-cycle plasma process uses noble gas (e.g. argon) as a working fluid. The noble gas is seeded with easily ionized material such as cesium. The closed-cycle plasma process is shown in Figure 3.

The closed-cycle liquid metal process is very similar to the closed-cycle plasma MHD, except that a gas liquid metal bath is used as a working fluid instead of the noble gas. Because liquid metal systems have high electrical conductivities compared to total gas systems, lower temperatures and magnetic fields are possible. This could result in smaller plant and higher extraction efficiencies. Figure 4 shows the closed-cycle liquid metal process.


REGENERATIVE HEAT EXCHANGER ARRAY
COAL
POWER
COAL TREAT-MENT
HOT GAS
AIR
BLOWER
AIR PREHEAT
PARTICULATE SEPARATION
COMBUSTOR
REHEAT PHASE
FLUE GAS PURGE PHASE
INERT GAS PURGE PHASE
BLOW DOWN PHASE
TO STACK
VACUUM PUMP
HEAT REJECTION
Heat Rejection
VACUUM PUMP
BLOWER
ARGON
ARGON PURIFICATION
POWER
COMPRESSOR
ARGON
MAGNET POWER
INVERTERS
AC POWER
CESIUM INJECTION
MHD GEN
DIFFUSER
CESIUM PURIFI-CATION
POWER
CESIUM
CS PUMP
SHAFT POWER
STEAM BOTTOMING PLANT
WASTE HEAT
COOLING TOWERS
POWER
WATER
PRECOOLER
HEAT REJECTION

SOURCE: Jackson 1976.

FIGURE 3
CLOSED-CYCLE PLASMA MHD


INVERTERS
AC POWER
MAGNET POWER
MIXER
GENERATOR
LM PUMP
SHAFT POWER
STEAM BOTTOMING PLANT
WASTE HEAT
COOLING TOWERS
POWER
HELIUM
STEAM HEAT
AFB FURNACE
TO STACK
HEAT REJECTION
PRECOOLER
SERIES OF MHD GENERATORS
LOW TEMP AIR PREHEATER
AIR ELECTRO STATIC PRECIPI- TATOR
HELIUM
SODIUM
COMPRESSOR
TO COAL TREAT MENT
SEPARATOR
DIFFUSER
LIMESTONE
AIR
COAL
COAL TREATMENT
EXHAUST HEAT
POWER
REJECTED SOLIDS
NOZZLE
HELIUM

SOURCE: Jackson 1976.

FIGURE 4
CLOSED-CYCLE LIQUID METAL MHD

Presently, the goal is to continue performing feasibility studies of the generator, heat exchanger, and materials in these processes. Their development not as advanced as the open-cycle MHD pending resolution of basic issues that present studies are addressing.

This joint U.S./USSR publication on MHD by Argonne National Laboratory outlines how this advanced process greatly reduces emissions of both SO2 and NOx.

Chapter 5

PROTECTION OF THE BIOSPHERE

5.1 INTRODUCTION

The combustion of fossil fuels in the generation of electricity degrades the environment through the emission of sulfur oxides, nitrogen oxides, and particulates to the atmosphere, and through the discharge of hot water to neighboring streams and lakes. Understandable concern for the environment has resulted in legislation severely restricting such emissions. Although various scrubbing systems have been investigated for control of sulfur oxides, the most controversial emissions, the utilities have been hesitant to adopt the suggested corrective techniques because of the high capital costs, the technological difficulties and consequent reduced availability of the scrubbing equipment, and the burdensome sludge disposal problems. The magnetohydrodynamic generation of electricity, although burning fossil fuels, offers the potential of completely eliminating the emission of sulfur oxides and substantially reducing emissions of nitrogen oxides and particulates, as well as decreasing thermal pollution.

5.2 SULFUR OXIDES

5.2.1 Removal of Sulfur Oxides

The most impressive advantage of open-cycle MHD power generation is its self-contained sulfur-removal system. A potassium salt is injected upstream of the MHD channel to provide the plasma with the electrical conductivity necessary for the generation of power. If this potassium salt seed is sulfur-free, it can react with and remove the sulfur oxides produced by fuel combustion.

Utilization of K_2CO_3 as seed for this purpose was studied in the pilot plant shown in Fig. 5.1. An atmospheric horizontal cyclone combustor was used, operating at a firing rate of 90 kg coal/h, a slag rejection of 75%, and an exit temperature of 2500 K. The combustion gas was cooled by flowing through three stages of radiant and one stage of convective heat exchange before entering the baghouse at 495 K. Practically complete (> 99.5%) recovery of the spent seed-fly ash mixture was obtained. When K_2CO_3 was added as seed at a concentration of one gram mole per kilogram of coal containing 2.2% sulfur, 99.8% of the SO_2 was removed.[1,2] Only 5 ppm of SO_2 remained in the gaseous effluent (see Table 5.1). This level of SO_2 is the same as obtained in the combustion of natural gas containing trace quantities of sulfur compounds. No contemporary flue gas scrubbing process re-

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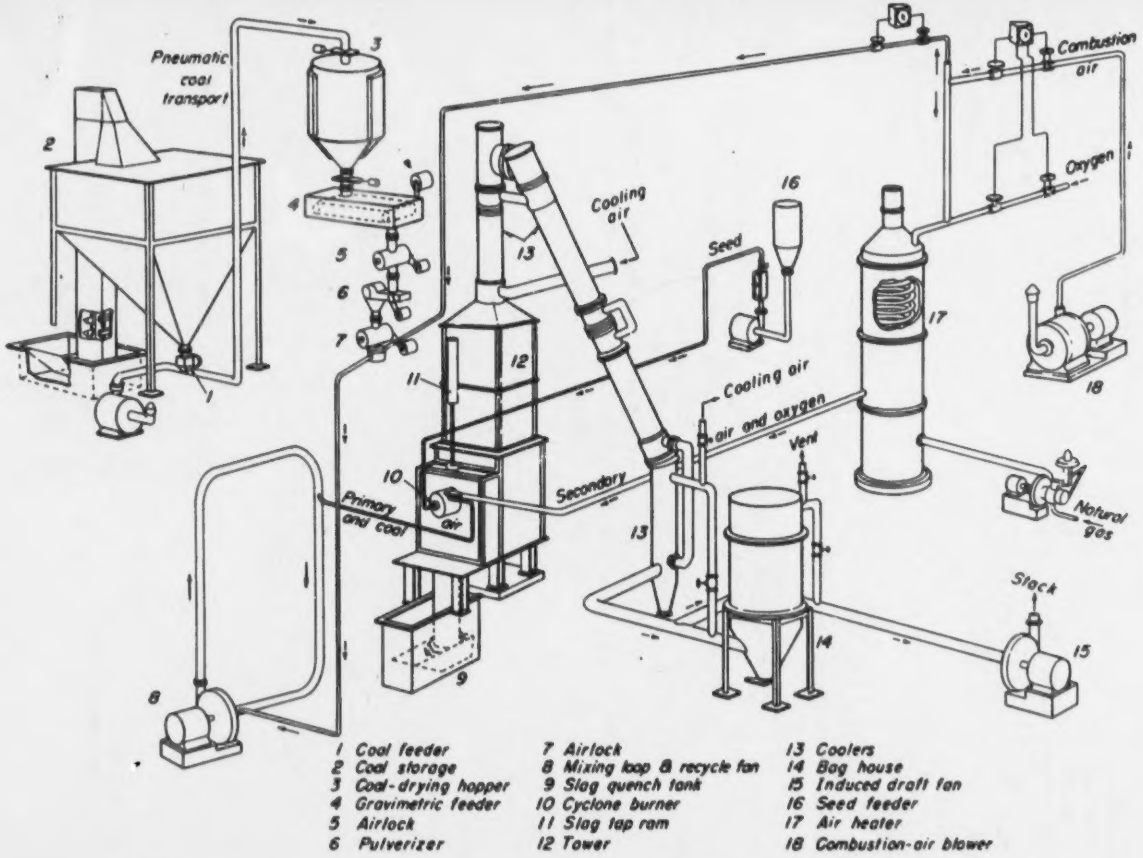

Pneumatic coal transport
Combustion air
Oxygen
Cooling air
Seed
Cooling air
air and oxygen
Vent
Secondary
Primary and coal
air
Natural gas
Stack
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16
17
18
1 Coal feeder
2 Coal storage
3 Coal-drying hopper
4 Gravimetric feeder
5 Airlock
6 Pulverizer
7 Airlock
8 Mixing loop & recycle fan
9 Slag quench tank
10 Cyclone burner
11 Slag tap ram
12 Tower
13 Coolers
14 Bag house
15 Induced draft fan
16 Seed feeder
17 Air heater
18 Combustion-air blower

Fig. 5.1—Experimental unit for studying the high-temperature combustion of coal for MHD power generation[1]

Table 5.1 Removal of Sulfur Oxides in MHD Combustion with K_2CO_3 as seed[2]

Mass % of sulfur in coal: 2.2 105% of stoichiometric oxygen $N_2/O_2 = 2$			
Seed Concentration $\frac{\text{g·mol } K_2CO_3}{\text{kg coal}}$	SO_2 in combustion gas, ppm	SO_2 removal, %	$\frac{\text{Actual } SO_2 \text{ removal}}{\text{Stoichiometric removal}^{(a)}} \times 100$
0.00	2735	—	—
0.37	1608	41.3	94
0.72	102	96.3	95
0.90	35	98.7	82
1.02	5	99.8	65

(a) Based on K_2SO_4 and partial loss of seed in slag.

moves SO_2 as effectively.

In future commercial MHD power plants operating at seed levels to 1.5 g-mol K_2O/kg coal, virtually complete removal of sulfur oxides from coal containing 4% sulfur is possible (see Fig. 5.2).[1] Thus, at the usual MHD seeding levels, SO_2 removal will meet the most stringent air pollution regulations and will permit the use of high-sulfur coals that are presently considered environmentally unsuitable for combustion in conventional power plants.

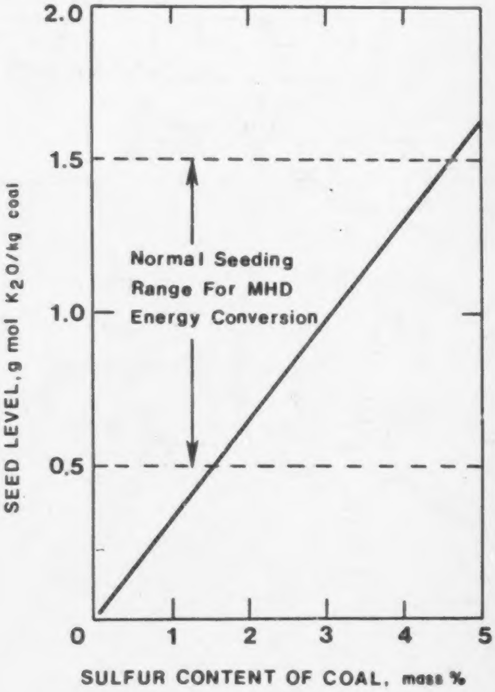


Fig. 5.2—Plasma seeding levels in removing sulfur oxide in coal combustion[1]

5.2.2 Seed Regeneration

To develop a continuous SO_2-removal system, the spent seed must be recovered and regenerated to remove the sulfur content before the seed is returned to the combustor. The MHD seed regeneration-desulfurization scheme is one of the most critical and least-researched areas of MHD. (The chemistry of various desulfurization schemes is outlined in Chapter 12, Ionizing Seed.) Earlier flow sheets and cost estimates for the Pittsburgh Energy Research Center (PERC) seed regeneration system[3] have been revised to reflect improved kinetic data,[4] the effects of inflation, and suggested process improvements. A 3000-MWt MHD power plant, fueled on 3 wt % sulfur coal, was selected as a base case for detailed analysis.

The flow sheet of the desulfurization process, shown in Fig. 5.3, essentially consists of two sections: a seed recovery section, in which the spent seed is separated from fly ash, and a seed desulfurization section, in which enough potassium sulfate is regenerated to potassium carbonate to remove all the sulfur oxides contained in the coal combustion products. The seed desulfurization is accomplished in two stages:

$$\text{reduction, } K_2SO_4 + xCO + (4\text{-}x)H_2 \rightarrow K_2S + (4\text{-}x)H_2O + xCO_2 \quad (5.1)$$

(x can vary between 0 and 4)

$$\text{and regeneration, } K_2S + CO_2 + H_2O \rightarrow K_2CO_3 + H_2S\uparrow . \quad (5.2)$$

The order of desulfurization and seed separation may possibly be reversed, and this option is presently under study at PERC. Of course, variations are possible; there is some indication, for example, that the desulfurization reaction can be conducted in a single step. Tables 5.2 and 5.3 list the chemical compositions and flow rates of the major solid, liquid, and gaseous streams involved. The fundamental assumptions for the base case studied are as follows:

1. The MHD plant had an efficiency of 50% and operated on Pittsburgh seam bituminous coal, which contains 10% ash and 3% sulfur and has a heating value of 32 MJ/kg. The coal flow rate into the MHD combustor is 95 kg/s.

2. The MHD seeding level is taken as 1 g· mol K_2O/kg coal (equivalent to 0.7 wt % K in the gas), and seed loss as 2%. This potassium

Table 5.2 Materials Balance for MHD Seed Desulfurization — Solid and Liquid Streams

	Solid and Liquid Streams Flow Rate, kg/h[a]							
A	K_2SO_4	58 997	Ash	3420				
B	Water	245 391						
C	K_2SO_4	589	Ash	3386				
D	K_2SO_4	58 380	Ash	34				
E	Coal	17 856						
F	Ash	1607						
G	K_2S	36 930	K_2SO_4	584	Ash	213		
H	Steam	3818						
I	K_2CO_3	44 927	K_2SO_4	584	K_2S	1108	213	
J	Sulfur	9886						
K	K_2SO_4	1186						

[a] See Figure 5.3 for stream identification

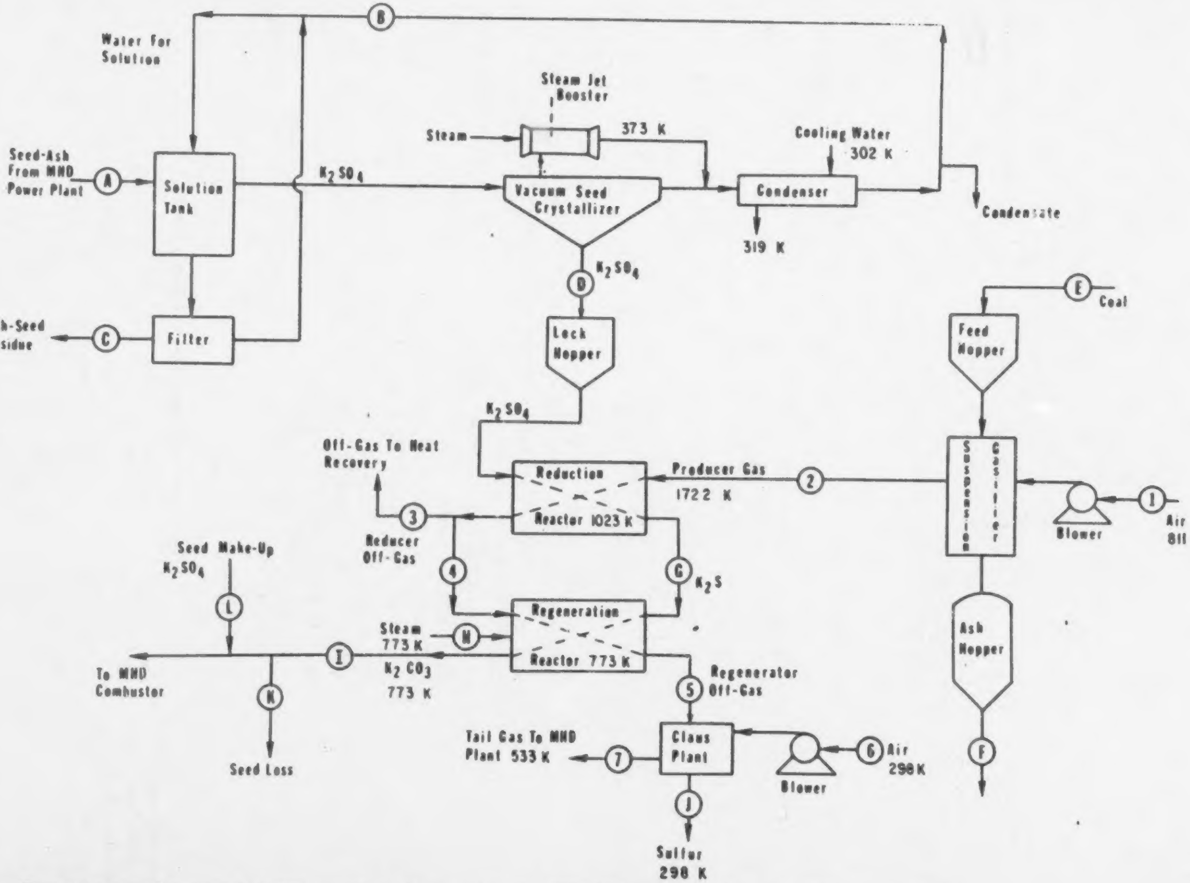

B
Water For Solution
Steam Jet Booster
Steam
373 K
Cooling Water
302 K
Seed-Ash From MHD Power Plant
A
Solution Tank
K2 SO4
Vacuum Seed Crystallizer
Condenser
Condensate
319 K
K2 SO4
D
E
Coal
Ash-Seed Residue
C
Filter
Lock Hopper
Feed Hopper
K2 SO4
Off-Gas To Heat Recovery
Reduction Reactor 1023 K
Producer Gas
1722 K
2
Suspension Gasifier
Blower
1
Air 811 K
3
Reducer Off-Gas
Seed Make-Up
K2 SO4
4
G
K2 S
L
Regeneration Reactor 773 K
Steam 773 K
H
To MHD Combustor
I
K2 CO3
773 K
Ash Hopper
K
Seed Loss
5
Regenerator Off-Gas
Tail Gas To MHD Plant 533 K
7
Claus Plant
Blower
6
Air 298 K
F
J
Sulfur 298 K

Fig. 5.3—Flow diagram of proposed MHD desulfurization process for 3000-MWt combined-cycle MHD power plant

seeding level is theoretically sufficient to eliminate all of the sulfur oxides produced by combustion of a 3.2 wt % sulfur coal. Existing Environmental Protection Agency (EPA) standards of 1.2 lb $SO_2/10^6$ Btu (5.14×10^{-10} kg SO_2/J), require only 74% sulfur removal. However, for this particular illustration, sufficient sulfur is removed (93.75%) to eliminate all the sulfur from a 3.0 wt % sulfur coal. The MHD seeding level can be adjusted upwards if this seed concentration does not prove adequate.

3. Hot water extraction of the spent MHD seed-fly ash mixture is 99% efficient, provided slag rejection in the MHD combustion system is 90%.

4. A CO-H_2 producer gas is obtained by operation of an air-blown suspension gasifier at 2×10^5 Pa pressure. Complete carbon conversion can be achieved with no tar carryover. Some fly ash carryover (less than 10%) will occur but will not affect operation of the seed regeneration system. The suspension gasifier is run at 50% of stoichiometric air, with an 800 K air preheat temperature and a 2.5% heat loss.

5. A 15% excess of producer gas is used to reduce the spent seed, to insure a reasonable reactor volume, in accordance with Eq. 5.1.

6. A 20% excess of reducer off-gas is employed for the desulfurization reactor, and a 1:1 ratio of CO_2 to H_2O is maintained by steam addition to the reactor, in accordance with Eq. 5.2.

7. A 99% conversion of K_2SO_4 to K_2S is assumed for the reduction reaction, and 97% conversion of K_2S to H_2S for the desulfurization reaction.

8. The H_2S level of 25.4% in the regenerator off-gas is considered sufficient to obviate the use of a liquid absorption system for gas purification. The regenerator off-gas can be sent directly to a Claus plant for the production of elemental sulfur.

9. A 96% conversion of hydrogen sulfide to elemental sulfur occurs in the Claus plant. The Claus plant tail gases may be recycled to the MHD exhaust gas stream, where any remaining H_2S or SO_2 will be captured as K_2SO_4.

Table 5.3 Materials Balance for MHD Desulfurization — Gaseous Streams

Number: Name: Composition, %	(1) Gasifier Air	(2) Producer Gas	(3) Reducer Off-gas to Heat Recovery	(4) Reducer Off-gas to Regeneration	(5) Regenerator Off-gas	(6) Claus Plant Air	(7) Claus Plant Tail Gas
A	0.9	0.7	0.7	0.7	0.7	0.9	0.8
CO_2	—	2.8	26.6	26.6	5.0	—	3.2
CO	—	25.6	1.9	1.9	0.8	—	—
H_2	—	9.1	1.2	1.2	2.0	—	1.3
H_2O	1.3	3.5	11.3	11.3	2.9	1.3	19.8
H_2S	—	0.3	0.3	0.3	25.4	—	0.7
N_2	77.1	57.9	58.0	58.0	63.1	77.1	75.2
O_2	20.7	—	—	—	—	20.7	—
Minor Components	—	0.1	1.0	1.0	0.1	—	0.3
Flow Rate, $m^3/h \times 10^{-3}$	68.00	90.56	59.60	29.62	28.45	18.41	42.25

Tables 5.4 and 5.5 provide equipment and capital investment data for the system under consideration. A capital cost of $26.60/kW was projected for this scheme, with the seed separation portion of the plant constituting the more expensive area. Estimated operating costs for the seed recovery and sulfur removal scheme are given in Table 5.6. An operating cost of 1.21 mills per kilowatt hour was computed. This appears a rather minor expense with reference to both fuel costs and capital charges, when viewed in the context of a central-station MHD power plant.

Table 5.4
Equipment Cost Summary for PERC Seed Regeneration-Desulfurization Scheme, 300 MWt, 1500 MWe

Liquids Processing	**Cost, $**
Evaporator	2 276 600
Circulating Pump and Drive	650 000
Vacuum Pump and Drive	162 500
Other	134 200
Direct Equipment Cost	3 223 300
Foundations, Electrical Instrumentation, Piping, etc.	3 384 300
Total Indirect (Overhead, Administration, Engineering, etc.)	4 498 700
Total Capital Cost	11 106 300
Solids Processing	
Reduction Reactor and Heat Exchanger	334 700
Regeneration Reactor and Heat Exchanger	246 000
Other	114 500
Direct Equipment Cost	695 200
Foundations, Electrical Instrumentation, Piping, etc.	747 700
Total Indirect	996 300
Total Capital Cost	2 439 200
Reducing Gas Production	
Cyclones	1 618 900
Gasifier	502 100
Gasifier Compressor	600 000
Other	129 300
Direct Equipment Cost	2 850 300
Foundations, Electrical Instrumentation, Piping, etc.	3 100 200
Total Indirect	4 924 800
Total Capital Cost	10 875 300
Claus Plant	
Total Capital Cost	2 250 000

Table 5.5
Total Capital Investment for PERC Seed Regeneration-Desulfurization Scheme, 3000 MWt, 1500 MWe

Item	**Cost, $**
Liquids Processing	11 106 300
Solids Processing	2 439 200
Reducing Gas Production	10 875 300
Claus Plant	2 250 000
Plant Facilities	2 000 300
Plant Utilities	2 867 100
Total Construction	31 538 500
Interest During Construction	4 730 800
Subtotal for Depreciation	36 269 300
Working Capital	3 626 900
Total Investment	39 896 400
Capital cost/kWe $\frac{\$39\,896\,400}{1\,500\,000} = \$26.60$	

Table 5.6 Estimated Annual Operating Cost of PERC seed Regeneration-Desulfurization Scheme, 3000 MWt, 1500 MWe

		Cost, $
Direct Cost		
Raw Materials and Utilities		
Coal — 17.83 t/h × $16.53/t × 7000 h/yr	2 063 100	
Power — 82.5 kW × $0.015/kW·h × 7000 h/yr	8 700	
Steam — 139 Mg/h × $1.32 Mg × 7000 h/yr	1 284 400	
Water — 79.5 m^3 × $0.04/$m^3$ × 7000 h/yr	22 300	
Make-up Seed — 1178.7 kg/h × $.22/kg × 7000 h/yr	1 815 200	
		5 193 700
Direct Labor		
192 man-hours/day × $6.00/man-hours × 365 day/yr	420 500	
Supervision (15% of labor)	63 100	
		483 600
Plant Maintenance		
39 men — $15,000/yr	585 000	
Supervision — 20% of maintenance labor	117 000	
Material — 150% of maintenance labor	887 500	
		1 579 500
Payroll overhead — 30% of payroll		355 700
Operating supplies — 20% of plant maintenance		315 900
Total Direct Cost		7 928 400
Indirect cost (administration and general overhead) 40% of labor, maintenance, and supplies		951 600
Total capital charges — 17% of total investment		6 782 400
Total Operating Cost		15 662 400
Credits		
Sulfur — 9.92 t/h × $27.65/t × 7000 h/yr		1 913 800
Steam — 2.86 MPa 700 K—64.84 Mg/h × $2.21/Mg × 7000 h/yr		1 003 100
Total Credits		2 916 900
Total operating cost after credits		12 745 500
Cost, mill/kW·h for 1500-MWe open-cycle MHD-steam power plant operating 7000 h/yr		1.21

5.3 NITROGEN OXIDES

5.3.1 NO_x Emissions on a National Scale

Nitrogen oxides (NO_x) are formed by natural processes and by the industrial activities of man. Both sources emit principally nitric oxide (NO). Although natural emissions, which occur as a result of bacterial and plant actions, exceed man-made emissions by a factor of 10 on a worldwide basis, the latter predominate in urban areas. Natural emissions of NO_x are considerable when the entire geographical area of the nation is considered, but because of their low-emission density they result only in an NO_2 background concentration of the order of $8 \mu g/m^3$. Man-made emissions of NO_x in the United States in 1972 amounted to 218×10^9 kg; Table 5.7 lists the sources and their contribution.[5]

5.3.2 National Standards for NO_x

Under the Clean Air Act of 1970, national ambient air-quality standards for nitrogen dioxide were established in the United States, as of April 30, 1971, at an annual average of $100 \mu g/m^3$ (0.05 ppm).

Table 5.7
Man-made NO_x Emissions in United States[5]

Source	Contribution, %
Stationary (60%)	
Utility	29.4
Industrial	10.8
Stationary internal combustion engine	11.4
Commercial residential space heating	4.2
Non-combustion sources[a]	4.2
Mobile (40%)	
Gasoline-powered motor vehicles	30.4
Diesel-powered vehicles	8.0
Railroads, aircraft, etc.	1.6

[a] e.g., nitric acid production, TNT production

On December 23, 1971, U.S. emission standards were set for new stationary fossil fuel-fired steam generating units of more than 250 million Btu/h (70 MJ/s) input for a two-hour average, expressed as NO_2 as follows:

gaseous fuels: 0.20 lb/million Btu (0.09 μg/J)
liquid fuels: 0.30 lb/million Btu (0.13 μg/J)
solid fuels: 0.70 lb/million Btu (0.30 μg/J)

In 1971, the U.S.S.R. Ministry of Health set maximum allowable concentrations of NO_x in the atmosphere[6] at 85 $\mu g/m^3$ for both instantaneous and 24-h average concentrations. However, in the presence of SO_x the maximum allowable concentration is decreased; the effects of SO_2 and NO_x are considered additive, and the following rule is applied:

$$\frac{(K_{SO_2})st}{K_{SO_2}} + \frac{(K_{NO_2})st}{K_{NO_2}} \geq 1,$$

where (K_{SO_2}) st and (K_{NO_2}) st are the maximum allowable concentrations and K_{SO_2} and K_{NO_2} are the measured concentrations of SO_2 and NO_2 in atmospheric air.

5.3.3 Control of NO_x

In fossil fuel combustion, nitrogen oxides are produced through fixation of the nitrogen in the air supplied to the combustor and, if oil or coal is used, from the nitrogen in the fuel as well. In coal combustion, approximately 50 to 80% of the NO in the exhaust gas is associated with the fuel nitrogen. Emissions of NO_x in coal firing tend to exceed those of gas and oil firing. An average of 0.3 μg NO_x/J for pulverized coal firing has been reported in a survey of vertical, corner, front wall, spreader stoker, and horizontally opposed combustion units.[7]

In MHD combustion, where temperatures are on the order of 3000 K (compared to 2000 K in conventional combustion), emissions of NO_x may be considerably greater. Some of the techniques used for NO_x control in conventional fossil fuel combustion are also applicable to NO_x control in MHD combustion, for example, 1) low-excess-air combustion, 2) staged combustion, and 3) flue gas treatment. Such other conventional techniques as flue gas recirculation, steam injection, and preheat reduction are not applicable to MHD combustion, because they result in lower combustion temperatures and, thus, lower electrical conductivities of the plasma.

Low-excess-air Combustion — This technique has been used successfully with all fuels, in a variety of furnaces, in conventional combustion. By reducing oxygen availability at the burners, both thermal and fuel NO levels are reduced. The degree of reduction is a function of furnace design, fuel characteristics, and excess air. Utilization of the low-excess air technique may be limited by the formation of incomplete combustion products, such as carbon monoxide, partially oxidized hydrocarbons, and carbon.

Staged Combustion or Off-stoichiometric Combustion — In conventional installations, two-stage combustion is practiced by operating the main burners at 90-95% of stoichiometric and adding the balance of the combustion air through separate ports above the main burner zone. A variation is to operate the lower set of burners in a fuel-rich mode and the upper burners fuel-lean.

Because the temperature of the combustion gas under fuel-rich condition is much higher in MHD combustion than in conventional combustion, air for completion of combustion must be added at a lower temperature, to avoid reformation of nitrogen oxides.

Flue Gas Treatment — The processes used to treat flue gases reduce NO_x emissions, either by decomposition of the NO_x to nitrogen and water or oxygen, or by removal of the NO_x from the gas stream. Dry systems are operated at approximately 600 K, wet systems at 310-325 K. Dry systems, which are further developed than the wet, use catalytic reduction, catalytic decomposition, and absorption. If the high-temperature region of the MHD process stream is maintained at fuel-rich condition, the oxygen deficiency causes the NO concentration to be lowered significantly. Figures 5.4a and 5.4b show the effect of various stoichiometries on equilibrium NO concentration as a function of temperature in the combustion of methane and coal.

This significant lowering of the NO results from the order-of-magnitude reduction in oxygen concentration produced by the change from air-rich to fuel-rich conditions. In the case of methane-air combustion products at 2000 K, for example, the equilibrium oxygen concentration for 110, 100, and 90% of the stoichiometric oxygen requirement is 1.66%, 0.078%, and 1.74 ppm, respectively. The corresponding numbers for coal-air combustion products are 1.82%, 0.100%, and 2.64 ppm.

At a location in the process stream where the gases have cooled to the point that the oxidation of nitrogen is no longer significant, the balance of the combustion air is added to oxidize the carbon monoxide. The resulting temperature after the balance of combustion air has been added must be low enough (<1600 K) that it is still below the level at which the oxidation of nitrogen occurs. The temperature at which to inject secondary air should be determined on the basis of detailed analysis of the thermal, flow, and stoichiometric conditions in the high-temperature zones of the systems.

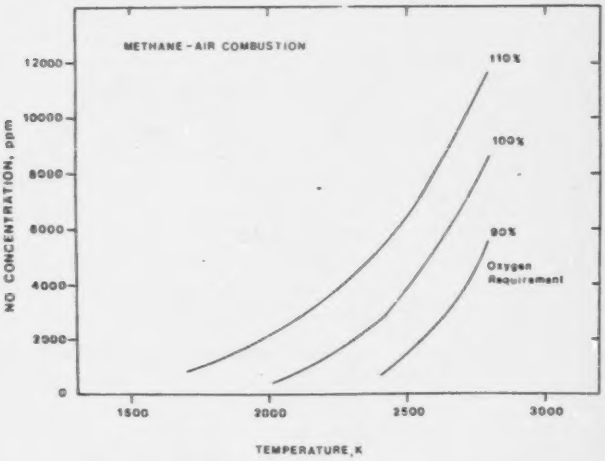
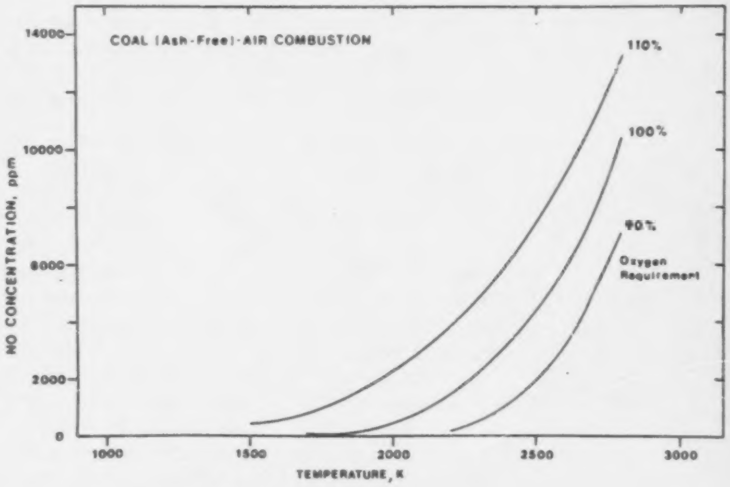


Fig. 5.4a—Equilibrium nitric oxide concentration in methane-air combustion gases

Fig. 5.4b—Equilibrium nitric oxide concentration in coal-air combustion gases



5.3.4 Experimental and Theoretical Studies of NO_x Reduction

Experimental Studies — Experimental data on NO_x control in coal combustion are shown in Table 5.8.[3] A cyclone combustor was operated at 88-110% of stoichiometric oxygen in the first stage. Ambient air was added at combustion gas temperatures of 1370 to 1840 K. With single-stage combustion and oxygen at 110% of stoichiometric, the concentration of NO_x leaving the stack was 4086 ppm. With two-stage combustion at high fuel-to-air ratios, the gas leaving the first-stage combustor had high H_2+CO content, so that the NO_x level rose with the increase in the second-stage combustion temperature when additional air was admitted. At lower fuel-to-air ratios, and as the temperature of the combustion gas was decreased at the location of additional

Table 5.8
NO_x Decomposition in Staged Combustion[3]

First-stage Exit Temperature, 2200-2260 K			Cooling Zone			Second Stage		
Stoich O_2, %	Effluent NO_x, ppm	Co, %	Residence Time, s	Final Temp, k	Calculated Final NO_x, ppm	Stoich O_2, %	NO_x, ppm	CO, %
110	4400	1.14	—	—	3815	110	4086	0
102	2695	2.17	—	—	2130	102	2612	0.1
101	2650	2.27	—	—	2056	101	2225	0.4
95	880	4.30	1.2	1370	680	106	150	0
94	782	6.61	1.1	1450	580	103	351	0
94	727	6.90	0.9	1620	580	105	550	0
92	543	8.11	0.9	1620	400	107	575	0
92	496	7.53	0.6	1840	400	107	885	0
91	496	9.95	0.6	1840	340	104	885	0
88	356	10.4	0.6	1840	230	104	814	0

air injection, NO_x emissions decreased. At 95% of stoichiometric air to the combustor and ambient air admitted at combustion gas temperature of 1370 K, NO_x emissions were only 150 ppm, or 0.052 μg NO_2/J input. In the two-stage combustion experiments, complete combustion of carbon monoxide was attained.

In these experiments, the measured NO concentrations were in good agreement with calculations of equilibrium NO concentrations at first-stage exit temperature of 2200-2260 K. This agreement indicates that the prior history of NO sources and details of NO generation are irrelevant. That is, the NO levels may have been higher for some upstream conditions because of higher temperatures or fuel-contained nitrogen; but the conditions and residence times were appropriate for decomposition of the NO to this (lower) equilibrium value.

As can be seen from the second and third lines in Table 5.8, operating in the low-excess-air regime reduces the NO levels, although a corresponding rise in carbon monoxide results. However, the NO concentration is not low enough to be acceptable in the low-excess-air situation. Only by operating the high-temperature zones fuel-rich was the NO lowered sufficiently to obtain allowable emissions.

In the experiments at Pittsburgh Energy Research Center, the temperature of the second-stage air was at ambient levels, and the air was injected at several locations. This had the effect of creating a gas cooling zone with various lengths and, thus, various residence times. Table 5.8 shows the residence times and final temperatures of the cooling zone.

Calculation of NO Decomposition Kinetics — Figure 5.5 shows the progress of decomposition of the NO as calculated for each of these cases with kinetic modeling. The model integrates the relevant chemical kinetic equations with the specified initial conditions appropriate for the experiment and the cooling rate of 750 K/s.

The equations are:

$$H+O_2=OH+O \tag{5.3}$$

$$O+H_2O=2OH \tag{5.4}$$

$$M+O_2=O+O \qquad (5.5)$$
$$CO+OH=CO_2+H \qquad (5.6)$$
$$O+NO=O_2+N \qquad (5.7)$$
$$N+NO=N_2+O \qquad (5.8)$$
$$H+NO=OH+N \qquad (5.9)$$

Equations 5.5, 5.7, and 5.8 comprise the Zeldovich mechanism. Equation 5.9 is important in reducing conditions and was added to improve the modeling accuracy for fuel-rich experiments. The reaction rate constants for the kinetic model calculations were taken from Baulch,[8] except for Eq. 5.9, where the data of McCollough[9] were used.

The results of the modeling are in reasonable agreement with the experimental measurements. In the fuel-rich experiments, reported in Table 5.8, the model shows that the NO decomposition in the cooling zone was, in some cases, to lower values than those measured in the effluent of the second stage. This indicates that additional NO can be generated in the second stage under certain conditions.

More importantly, however, in some cases the second-stage air addition can evidently provide additional opportunity for NO decomposition. Where the stoichiometry in the high-temperature region was 94-95% of required oxygen, NO in the second-stage effluent was lower

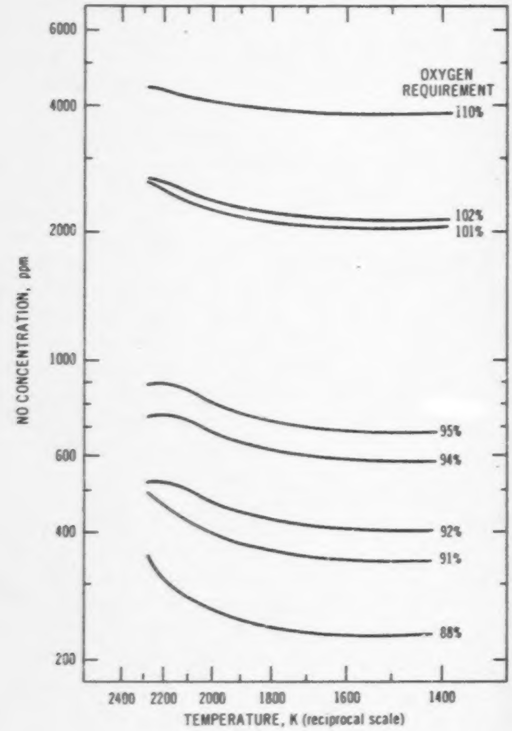


Fig. 5.5—Nitric oxide concentrations computed by kinetic modeling in cooling zone of experimental MHD coal combustion stand

than that computed from the kinetic model for the cooling zone. Apparently, after cooling to 1370-1450 K at the higher stoichiometry, the remaining concentration of combustibles was not sufficient to cause an increase in NO_x concentration. The reduction in NO_x levels below the computed values may be a consequence of the heterogeneous reduction of NO by CO and H_2. It has been demonstrated that such NO reduction reactions are important in the presence of alumina[10] and in the presence of char and ash in coal combustion products.[11] These reactions may reduce the levels of NO_x emissions below those calculated from the homogeneous reactions for coal combustion products, although not below the levels calculated for the ash-free combustion products of natural gas or oil. Available experimental data are insufficient, however, for detailed analysis of combustion in the second stage. Additional experimental data are required for better understanding of the mechanism of NO_x during low-temperature secondary combustion.

The last three columns of Table 5.8 show the measured effluent concentrations of NO and CO and the stoichiometric oxygen level in the second stage. For this experimental combustor, the optimum mode of operation for NO emission control was with 95% of the stoichiometric oxygen requirement in the first stage, followed by residence in a cooling zone until gas temperature dropped to 1400 K, and, finally, addition of air to 106% of the stoichiometric requirement.

5.3.5 Extension of Two-stage Operation to Large-scale MHD Plants

Kinetic calculations indicate it is possible to reduce the concentration of NO_x in large-scale MHD power plant effluent by operating the combustor fuel-rich and supplying sufficient air downstream of the MHD generator to obtain complete oxidation.[3,12-14] Good correlation was shown between experimental and analytical data in the combustion of natural gas. The six-reaction Lavoie model gives essentially the same NO profile as a more complex model using 60 homogeneous gas reactions.[15] In the range 2300-2000 K, the more important reaction in nitric oxide decomposition is $N+NO \rightarrow N_2+O$; in the 2000-1800 K range, the reaction $H+NO \rightarrow N+OH$ is more important. Further reduction of NO is possible at lower gas temperatures by means of catalytic decomposition over an alumina surface, where the decomposition through gas phase reactions is negligible.[16]

The kinetics of NO decomposition were calculated with the same kinetic model described in Eqs. 5.3-5.9, for a variety of cooling rates and stoichiometries starting with equilibrium conditions at 2400 K. The results are shown in Figs. 5.6a and 5.6b. In Fig. 5.6a, the composition was that for methane-air combustion at 110% 100%, and 90% of the stoichiometric oxygen requirement. As is obvious in these curves, the kinetics of NO decomposition can proceed toward equilibrium until some "low" temperature is attained. That temperature depends upon the cooling rate of the gas; that is, at higher cooling rates the kinetics cannot follow equilibrium for as long a time as at lower cooling rates. The cooling rates were chosen to include those that might reasonably be expected in a full-scale MHD power plant. Figure 5.6b shows the same curves for a coal-air combustion situation (the ash constituents were ignored in calculating the oxygen requirements). The curves for coal are qualitatively similar to those for methane, differing only in that the frozen NO_x levels are slightly higher. It should be noted, however, that these curves do not include the effect of any heterogeneous reactions. It is necessary, however, to compare the experimental conditions with those in a full-scale plant. Several such studies have been carried out.

The temperature-time characteristics for the experimental combustion facility and for a conceptual 1000-MW coal-fired MHD central station plant[17] are shown in Fig. 5.7. It was assumed that this central station model had an air-preheat temperature of 1475 K and a gas cooling rate of 200-400 K/s in the radiant boiler and 100 K/s in the air preheater. These cooling rates are significantly lower than the 750 K/s

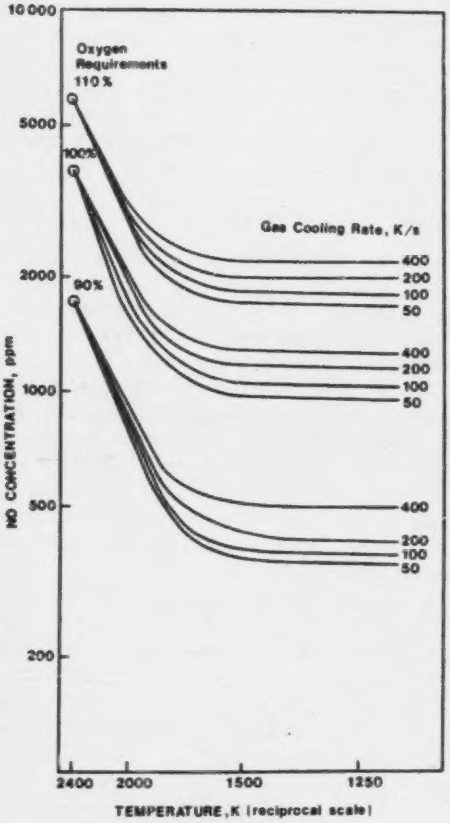


Fig. 5.6a—Nitric oxide concentrations computed by kinetic modeling for several cooling rates and stoichiometries for methane-air combustion gases

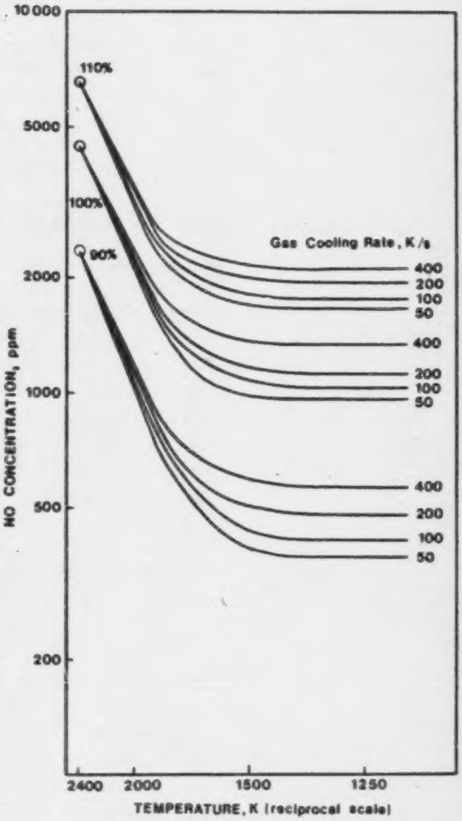


Fig. 5.6b—Nitric oxide concentrations computed by kinetic modeling for several cooling rates and stoichiometries for coal-air combustion gases

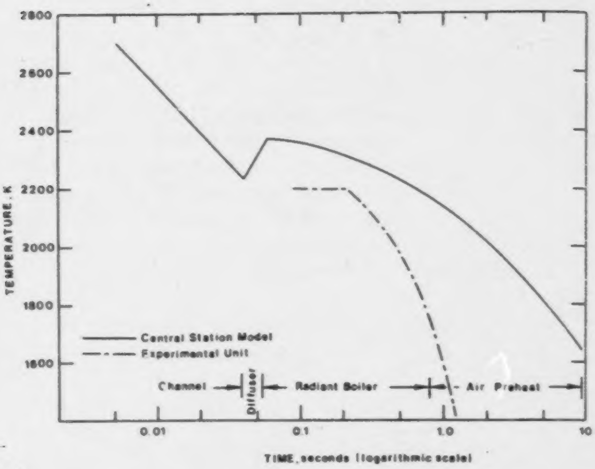


Fig. 5.7—Temperature-time profiles for the experimental MHD coal combustion stand and for the proposed coal-fired 1000-MW central-station MHD plant[17]

experienced in the experimental combustion facility. The results for coal combustion at 90% stoichiometric oxygen and a cooling rate of 200 K/s were plotted on the time profile for the plant downstream of the MHD generator. The resulting curve is shown in Fig. 5.8. For comparison, the equilibrium values of NO concentration are also shown. Although the kinetic-controlled NO levels are much above equilibrium, they are well within acceptable limits, which indicates that NO concentrations in full-scale MHD plants can be controlled by staged-air addition. The recommended point of air addition in a plant corresponding to this model would be near the exit of the air preheater.

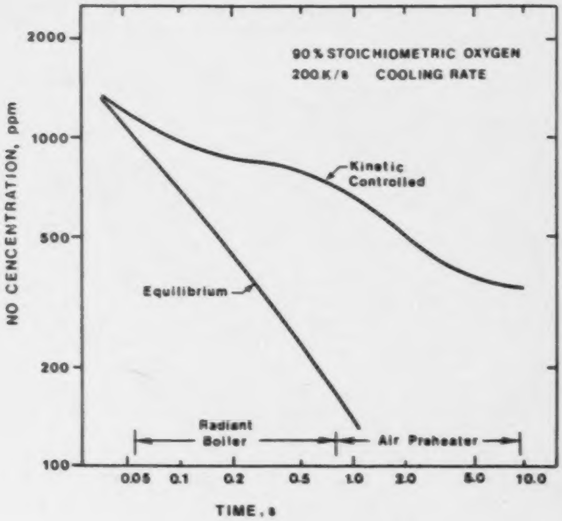


Fig. 5.8—Equilibrium and kinetic-controlled NO concentration in coal-fired central-station MHD plant

Similar results were obtained from a study on a natural gas-fired MHD plant carried out by the Institute of High Temperatures of the U.S.S.R. Academy of Sciences. Kinetically calculated curves of NO decomposition have been developed by the Institute and are shown in Figs. 5.9a and 5.9b.[18] The former is computed at 90% of the stoichiometric air requirement and the latter at 95% stoichiometry. In both cases, the residence time of combustion products in the radiant chamber is approximately three seconds in the temperature region of 2300-1600 K. As can be seen in these curves, the appropriate gas cooling rate and stoichiometry make it possible to obtain nitric oxides concentrations of 200 ppm in the flue gas of MHD power plants.

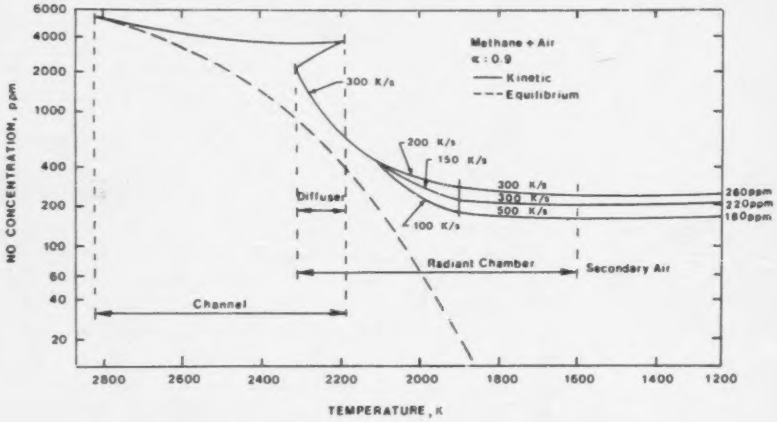
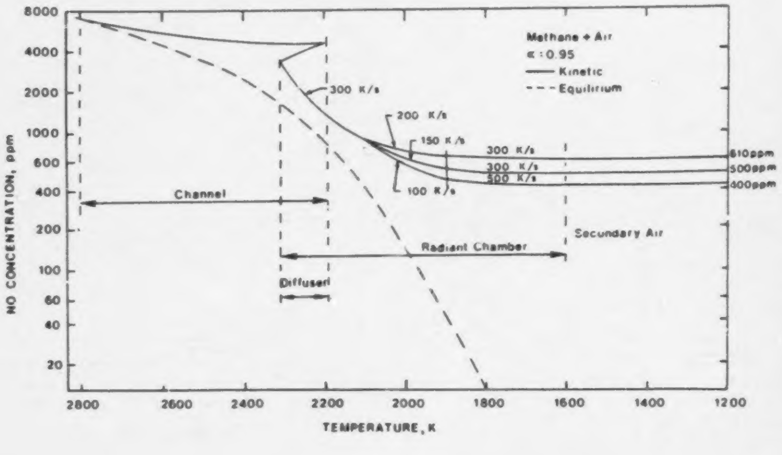


Fig. 5.9—Equilibrium and kinetic curves of NO content in the combustion products of a commercial MHD power plant fired with natural gas at (a), above, 90% stoichiometry and (b), below, 95% stoichiometry[18]



The flue gas cooling rate required in MHD plants needs to be explored in greater detail; in particular, optimization studies need to be carried out. For more complete NO decomposition, low flue gas cooling rates are necessary and, consequently, relatively long residence times in the heat-exchange apparatus; this, however, leads to correspondingly higher capital costs. Analyses have indicated that the optimal effect is achieved with a variable flue gas cooling rate. The minimal cooling rate should be in the range of 2100 to 1900 K, above and below which cooling may be more rapid. It is more efficient to have faster cooling below 1900 K, where the NO decomposition rate is reduced, rather than above 2100 K, where the equilibrium NO concentration is still rather large. At temperatures below 1600 K, NO seldom decomposes, except in the presence of a catalyst.

The combustion stream temperature at which the secondary air is injected must also be taken into consideration; it should be low enough to avoid the possible increase of NO_x emissions. However, second-stage burning of combustible species in the effluent at temperatures below 1600 K is not desirable, because 1) it is difficult to obtain complete combustion of CO at low temperatures in large-scale equipment, especially in view of the likelihood of poor mixing of the oxidizer and the gas, and 2) when incomplete combustion takes place in the presence of insufficient oxidizer, particularly in facilities utilizing liquid fuel or coal, the combustion products may contain carcinogenic compounds, usually polycyclic aromatic hydrocarbons such as benzopyrene ($C_{20}H_{12}$).

An increase in emission levels of benzopyrene, an unstable intermediate combustion product during coke and soot formation, has been noted in stack gases from commercial steam generators when near-stoichiometric levels of the oxidizer are used to suppress NO_x formation.[19] At the high temperatures of MHD-plant combustors, chemical reaction rates increase, and the possibility of residual intermediate combustion products is reduced. However, because of the stringent air

Table 5.9
Influence of Stoichiometric Air Ratio on Capital Investment — 2000 MWt Coal-fired MHD Power Plant[3]

Combustor pressure, MPa	0.8		
Plasma seeding level, g·mol K_2SO_4/kg coal	2.0		
Gas velocity in generator, m/s	900		
Magnetic field, T	6		
Maximum Hall parameter	4		
Loading parameter	0.8		
First-stage air preheat temperature, K	1470		
Second-stage air preheat temperature, K	300		
Overall stoichiometric air ratio	1.05		
Front end stoichiometric air ratio	0.87	0.95	1.05
Flame temperature, K	2792	2780	2760
MHD power, MW	558	582	618
Steam power, MW	421	403	364
Compressor power, MW	164	179	198
Overall efficiency, %	49.50	49.85	49.75
Relative generator length	1.192	1.000	1.305
Relative power density	0.870	1.000	0.713
Relative capital cost	1.019	1.000	1.098

pollution regulations concerning benzopyrene (e.g., U.S.S.R. standard, 0.1 $\mu g/m^3$ maximum), it must be completely eliminated in the second stage, which requires a higher level of excess oxidizer and relatively high temperatures.

Plant efficiency calculations and preliminary cost estimates have been done for an MHD plant incorporating two-stage combustion for NO_x emission control. Among the factors included were the reduced compression requirements and the enhanced electrical conductivity for the fuel-rich first-stage combustion. The results are shown in Table 5.9. Two-stage combustion with the first stage at 95% of stoichiometric air is approximately 10% lower in capital cost, with no loss in efficiency, compared to single-stage operation at 105% of stoichiometric air.[3]

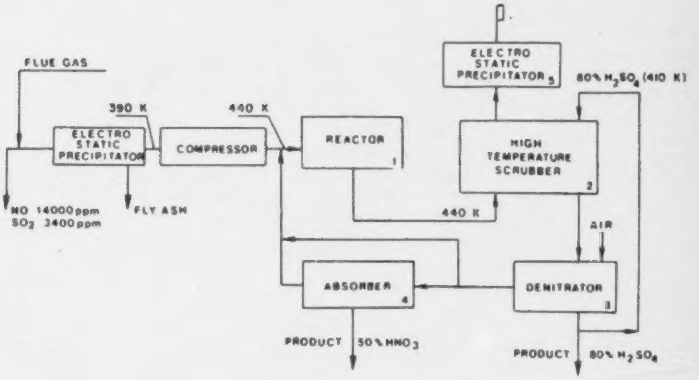
5.3.6 Flue Gas Scrubbing of NO_x

Several absorption systems for scrubbing NO_x have been proposed, but none has been developed to pilot plant stage.[20]

In contrast to the flue gas of conventional steam turbine power stations, MHD combustion products contain high nitric oxide concentrations, which improves the economics of flue gas cleaning. Nitric acid, which is a fertilizer material and has many industrial applications as well, can be produced as a by-product.

The promise of this approach is indicated by the modified Tyco sulfuric acid production process for simultaneous removal of sulfur and nitric oxides from flue gases.[21] The by-products of the process, nitric and sulfuric acids, substantially offset the cost of the chemical recovery system. The system is shown schematically in Fig. 5.10.

In this process, SO_2 in the flue gas is oxidized by NO_2 to H_2SO_4 in the reactor. The flue gas is then washed with sulfuric acid in the high-temperature scrubber, where equimolar ratio of NO_2 and NO is maintained in the gas by NO_2 drawn from the denitrator, and nitrosyl sulfuric acid is produced. The nitrosyl sulfuric acid is decomposed to


FLUE GAS
390 K
440 K
ELECTRO STATIC PRECIPITATOR
COMPRESSOR
REACTOR 1
ELECTRO STATIC PRECIPITATOR 5
80% H_2SO_4 (410 K)
HIGH TEMPERATURE SCRUBBER 2
NO 14000 ppm
SO_2 3400 ppm
FLY ASH
440 K
AIR
ABSORBER 4
DENITRATOR 3
PRODUCT 50% HNO_3
PRODUCT 80% H_2SO_4

1. $SO_2 + NO_2 + H_2O \quad H_2SO_4 + NO$
2. $2H_2SO_4 + N_2O_3 \quad 2HNSO_5 + H_2O$
3. $2HNSO_5 + O \; SO_2 + H_2O \quad 2H_2SO_4 + NO_2$
4. $3NO_2 + H_2O \quad 2HNO_3 + NO$
5. Mist Removal H_2SO_4

Fig. 5.10—Schematic of SO_2 and NO_x removal from plant effluent

80% H_2SO_4 and NO_2; the NO_2 is in part converted to 50% nitric acid and in part recirculated to the scrubber.

The scheme shown in Fig. 5.10 has certain deficiencies and is not optimal for cleaning the flue gases of an MHD electrical power station. However, the process may be used as a model for evaluating the technical-economic aspects of methods for the simultaneous removal of SO_2 and NO_x from MHD flue gases, because it has many industrial analogs. The data thus obtained are conservative and illustrate the promise of the approach.

5.3.7 Economics of Simultaneous NO_x and SO_2 Scrubbing

For comparison of technical-economic parameters, three different 3000-MW thermal electrical power stations were analyzed on the basis of using high-sulfur coal of the following composition:

Coal Analysis (Wt %)

Moisture	8
Ash	23
C	55.2
H	3.8
S	3.2
N	1.0
O	5.8
Heating value	22 MJ/kg

To meet environmental standards governing harmful emissions to the atmosphere, these stations are equipped with SO_2 and NO_x control systems.

Case No. 1: A steam-turbine power station with units of 500 MW with initial steam parameters of 24 MPa pressure and 813 K/813 K temperature.[22] Wet scrubbing of flue gas is used to decrease SO_2 emissions. Technical and economic parameters of such a system are assumed from the average parameters of "wet system" SO_2 removal from exhaust gases.[23]

Case No. 2: An MHD power station with units of 940 MW and steam turbine units of 500 MW. Air is preheated to 2270 K in Cowper stoves fired with natural gas;[24] coal is burned in two stages for NO_x control, and sulfur oxides are removed by seed.

Case No. 3: An MHD power station with units of 940 MW similar to Case No. 2. However, SO_2 and NO_x are removed simultaneously from the flue gas at the exit of the installation. In this case, the seed is not regenerated, and it is introduced into the combustor in the form of sulfate. The economics of this case improves with increase in NO_x removal. Concentration of NO_x is maintained at high levels, 80% of the equilibrium level in the combustor, through use of a supersonic channel and flue gas recycling. The costs of compressing the flue gases and maintaining high NO_x content in the flue gases through recirculation are taken into account. In Table 5.10 the cost of the system for SO_2 and NO_x removal, modeled after Fig. 5.10, is calculated by analogy with equipment costs for sulfuric acid production.

Table 5.10
Capital Expenditures for a System for SO_2 and NO_x Removal in a 940-MW MHD Power Plant (Flue gas volume= 1258×10^3 m^3/h)

Name	Cost, rubles$\times 10^3$
Reactor	150
High-Temperature Scrubber	5 100
Catalytic Nitrogen Remover	4 100
SO_2 Absorber	252
Electric Filters	1 280
End Fan	425
Equipment Cost	11 307
Equipment Assembly — 30%	3 400
Pipes — 25%	2 830
Control and Measuring Equipment — 8%	905
Building, Construction — 35%	5 960
Subtotal Cost	22 402
General Station Expenditures (30% of Subtotal Cost)	6 720
Total	29 122
Capital Cost (rubles/kW)	31

Table 5.11 lists basic capital cost data for all three cases. Table 5.12 gives technical-economic parameters of the compared cases, which indicate that the MHD power plant with two-stage combustion (Case No. 2) has the economic edge over the steam-turbine electrical power station with flue gas cleaning, especially at higher coal costs. The MHD power plant with SO_2 and NO_x removal (Case No. 3) has a lower cost of electrical power production throughout the entire range of coal costs examined. It should be noted that, although the version incorporating simultaneous SO_2 and NO_x removal is promising, it requires further

Table 5.11
Basic Capital Cost Data for 3000-MWt MHD and Steam-Turbine Plants Equipped with SO_2 and NO_x Control Systems

Description	Capital Cost rubles/kW	$/kW	%
1. Capital expenditures for steam-turbine power plant of 3000 MW with 500 MW units burning coal	132[22]	190	100
2. Same as above with natural gas firing	120[22]	172	91
3. Capital expenditures for an MHD power plant of 3000 MW with 1000-MW units operating on natural gas	150-130[24]	214-185	113-98
4. Coal-processing plant for a 3000-MW MHD plant	6	8.5	4.6
5. Boiler cost increase when 25% of stack gases are recycled	1	1.43	0.8
6. Boiler cost increase due to radiation chamber for NO_x decomposition	3	4.3	2.3
7. Capital expenditures for seed separation system	1.6	2.3[3]	1.2
8. Capital expenditures for seed recovery (regeneration) system	7.4	10.65[3]	5.5
9. Capital expenditures for stack gas SO_2 scrubbing	—	—	15-25[23]
10. Capital expenditures for the SO_2 and NO_x removal system	31	43	23.5

Table 5.12
Economic Comparison of SO_2 and NO_x Control in MHD and Steam-Turbine Power Plants

Index	Steam Power Plant Equipped with a System of Purifying Exhaust Gases	MHD Power Plant	
		Two-stage Combustion	SO_2 and NO_x Capture from Exhaust Gases
	I	II	III
1. Power plant electric power, MW	3000	3000	3000
2. Fuel			
Coal (S=3.2%), %	100	59.8	59.8
Natural gas, %		40.2	40.2
3. Hours operated per year	6500	6500	6500
4. SO_2 and NO_x content in combustion products downstream from the MHD channel			
SO_2, ppm	3400	3400	3400
NO_x, ppm	800-1000	3000	14000
5. SO_2 and NO_x content at the end of the stack			
SO_2, ppm	<300	<100	<300
NO_x, ppm	300-400	<300	<200
6. Efficiency, %	38.5	48.5	46.9
(for its own operation), %	6.5	4.5	7.1
7. Capital expenditure of power plant, %	100	118-103	118-103
(rubles/kW)	(132)	(156-136)	(156-136)
8. Expenditures to protect the surrounding medium			
a. Capital expenditures			
—Cleaning the exhaust gases from SO_x, %	15-25	—	23.5
(rubles/kW)	(20-33)		(31)
—Separation and regeneration of seed, %	—	6.7	1.2
(rubles/kW)	—	(8.9)	(1.6)
—Increase in boiler cost, %	—	2.3	0.8
(rubles/kW)	—	(3)	(1)
b. Operating cost at 22.6% of capital expenditures annually			
Kop/kW·h	0.2-0.3	0.06	0.15
(mill/kW·h)	(2.6-4.0)	(0.9)	(2.0)
9. Production of H_2SO_4 and HNO_3 (g/kW·h)			
80% H_2SO_4	47	22	23
50% HNO_3	—	—	97
10. Cost of production			
80% H_2SO_4 (at 21 rubles per ton)			
Kop/kW·h	0.1	0.046	0.048
(mill/kW·h)	(1.43)	(0.66)	(0.68)

Table 5.12 cont'd.

Index	Steam Power Plant Equipped with a System of Purifying Exhaust Gases	MHD Power Plant	
		Two-stage Combustion	SO_2 and NO_x Capture from Exhaust Gases
	I	II	III
50% HNO_3 (at 23 rubles per ton)			
Kop/kW·h	—	—	0.22
(mill/kW·h)			(3.14)
11. Cost of natural gas at a cost of 24 rubles ($34.4) for 29 GJ			
Kop/kW·h	—	0.244	0.252
(mill/kW·h)	—	(3.2)	(3.3)
12. Part of capital expenditures per cost of electric power assuming 22.6% per year			
Kop/kW·h	0.49	0.57-0.495	0.583-0.509
(mill/kW·h)	(7.0)	(8.1-7.1)	(8.3-7.2)
13. Salary			
Kop/kW·h	0.012	0.012	0.012
(mill/kW·h)	(0.17)	(0.17)	(0.17)
14. Cost of coal assuming:			
C=7 rubles ($9.2) per 29 GJ			
Kop/kW·h	0.224	0.107	0.11
(mill/kW·h)	(3.2)	(1.53)	(1.57)
C=10 rubles ($13.2) per 29 GJ			
Kop/kW·h	0.32	0.152	0.157
(mill/kW·h)	(4.58)	(2.17)	(2.24)
C=15 rubles ($19.8) per 29 GJ			
Kop/kW·h	0.48	0.228	0.235
(mill/kW·h)	(6.86)	(3.26)	(3.36)
C=20 rubles ($26.4) per 29 GJ			
Kop/kW·h	0.64	0.304	0.314
(mill/kW·h)	(9.15)	(4.35)	(4.49)
15. Cost of electric power assuming:			
C=7 rubles ($9.2) per 29 GJ			
Kop/kW·h	0.82-0.92	0.87-0.95	0.76-0.84
(mill/kW·h)	(11.7-13.2)	(12.4-13.6)	(10.9-12.0)
C=10 rubles ($13.2) per 29 GJ			
Kop/kW·h	0.92-1.02	0.92-1.07	0.81-0.89
(mill/kW·h)	(13.2-14.6)	(13.1-14.2)	(11.5-12.7)
C=15 rubles ($19.8) per 29 GJ			
Kop/kW·h	1.08-1.18	0.09-1.07	0.89-0.96
(mill/kW·h)	(15.5-16.9)	(14.2-15.3)	(12.7-13.7)

Table 5.12 cont'd.

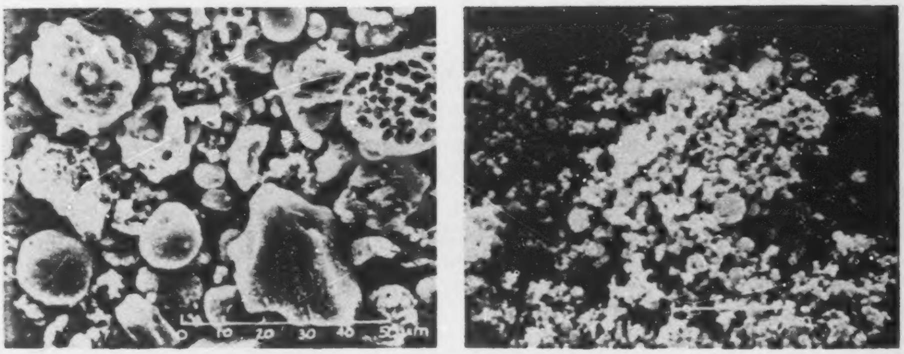
Index	Steam Power Plant Equipped with a System of Purifying Exhaust Gases	MHD Power Plant	
		Two-stage Combustion	SO_2 and NO_x Capture from Exhaust Gases
	I	II	III
C=20 rubles ($26.4) per 29 GJ			
Kop/kW·h	1.24-1.34	1.071-1.07	0.97-1.04
(mill/kW·h)	(17.8-19.2)	(15.3-16.3)	(13.9-14.9)
16. Savings in terms of cost in electric power in % assuming:			
C=7 rubles ($9.2) per 29 GJ	0	(−6.1)—(−3.2)	(+7.3)—(+8.7)
C=10 rubles ($13.2) per 29 GJ	0	(0.0)—(+2.9)	(+11.3)—(+12.6)
C=15 rubles ($19.8) per 29 GJ	0	(−8.3)—(+9.3)	(+17.6)—(+18.7)
C=20 rubles ($26.4) per 29 GJ	0	(+13.7)—(+14.9)	(+21.7)—(+22.4)

experimental investigation, whereas the version with NO_x decomposition (Case No. 2) is more fully developed and is considered ready for industrial application.

5.4 PARTICULATES

The particulates in the flue gas of an MHD plant before their removal by control equipment such as electrostatic precipitators, Venturi scrubbers, or baghouse fabric filters consist of seed and mineral matter in the fuel. For a 2000-MWt, coal-burning MHD plant using a 3.2% sulfur coal, with 9% ash and 20% ash carryover, K_2SO_4 will be emitted in the flue gas at the rate of 11 kg/s, and ash at the rate of 1.4 kg/s.[17] The size distribution in the MHD plant will not be the same as in a conventional fossil-fuel plant. In conventional pulverized-coal power plants, the size of the fly ash is correlated with the coal particle size and, typically, has a median diameter of 20-50 μm. In an MHD plant, however, the high temperature in the MHD combustor will melt or vaporize the ash components, and the size distribution of the resulting fly ash will be controlled by the physical phenomena of condensation and agglomeration of ash and seed in the combustion gas.

Preliminary evidence shows that the size distribution of the fly ash in an MHD plant will be an order of magnitude smaller than that of fly ash from a conventional power plant. Scanning electron micrographs of fly ash obtained in a high-temperature pilot plant and in conventional pulverized-coal combustion are compared in Fig. 5.11.[3] Because of the significant percentage of submicrometer-size particles in the MHD effluent and the need to recover the water-soluble seed, it may be advantageous to have a Venturi scrubber for removing the larger ash particles and trapping the water-soluble seed and a fabric filter downstream of the Venturi scrubber to remove the finer ash particles.



Conventional High-temperature

Fig. 5.11—Fly ash distribution in conventional pulverized coal combustion and in a high-temperature pilot plant[3] coal combustion

5.5 THERMAL POLLUTION

Because electric generating plants comprise by far the largest industrial user of cooling water (accounting for 80% of the total), thermal pollution will increase sharply with the expected tremendous growth of this industry. With increase in power plant efficiency, however, less waste heat is discharged to the cooling water.[25] Figure 5.12 shows the effect of efficiency on thermal pollution. The most modern fossil-fuel plant at 40% efficiency releases to the cooling water 1.1 kW per kW produced. A present-day nuclear plant of 33% efficiency releases 1.88 kW—a significantly higher degree of thermal pollution. Because the number of nuclear power plants is expected to increase with time, heat rejection for the predicted mixture of nuclear and fossil plants is expected to increase ninefold in the United States by the year 2000.[26] MHD plants,

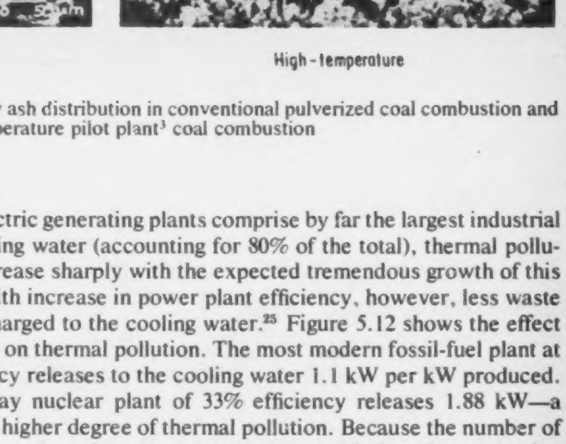


Fig. 5.12—Thermal pollution as a function of power plant efficiency[25]

with their considerably higher efficiency, could help stem this rise in thermal pollution. A first-generation MHD plant of 52% efficiency rejects to its cooling water 0.82 kW per kilowatt produced, or 43% the rate of the nuclear plant.

5.6 EFFECT ON CORROSION OF CONTROLLING STOICHIOMETRIC CONDITIONS

Fireside corrosion and fouling of furnace, boiler, and superheater tubes in coal- and oil-fired boilers have been recognized for many years. Extensive research has been conducted into the mechanisms of the reactions, and it has been determined that alkali metals and sulfur in the fuel are the precursors of the problem. These problems are discussed more fully in Chapters 11 and 13. A brief discussion of the potential impact of controlling the stoichiometry to achieve environmental control is presented here, however, to underscore the importance of the technical area and the interactions between the processes cited and plant operation and materials corrosion.

Alkali seeding to increase the ionization of the combustion gas in the MHD process can increase the alkali content of the effluent ten- to twenty-fold. To ascertain the problems that high-alkali concentrations would cause, preliminary tests were run with seeded and non-seeded coal combustion gases.[27] The gas leaving the combustor flowed through a heat exchanger consisting of air-cooled tubes fabricated of a variety of metals and alloys. The gas-side surface temperature of the tubes was controlled between 700 and 1090 K. The degree of corrosion was determined by the thinning of the tube walls, and the nature of the attack by metallographic examination of the tube metal and by analysis of the deposits found on the tubes. In a 100-h test, tube materials exposed to the seeded flue gas were more severely attacked than those exposed to the non-seeded flue gas. Haynes 25, a chromium-nickel alloy, was satisfactory for use at 1090 K wall temperature and 1645 K seeded flue gas temperature. No corrosion was experienced by 310, 316, and 446 stainless steel at 870 K wall temperature and 1425 K gas temperature. Carbon steel and Croloy 5 (a 5% chromium-carbon steel) were attacked at wall temperature of 700 K and flue gas temperature of 1255 K.

In combustion at excess-air conditions, corrosion can occur by removal of the protective scales on metal tubes through chemical reaction with a corrodent formed within an overlying deposit, by reaction of the protective scale with constituents in the flue gas and the overlying deposit, or by direct attack of the metal surface.[28] At excess-air conditions, sulfidation, in which the iron tubing is attacked by sulfur released from pyrites, is not a major cause of corrosion. When an oxidizing atmosphere is not available, however, as would be the case beneath a deposit, elemental sulfur can be released from pyrites and attack the tube metal. In two-stage combustion, in which the combustor is operated under reducing conditions and secondary air is added at lower temperatures to control nitric oxide formation, increased sulfidic corrosion is possible. An equilibrium plot of the distribution of the sulfur-containing compounds was computed on the basis of conditions described in subsection 5.3.4: 95% of stoichiometric oxygen to the

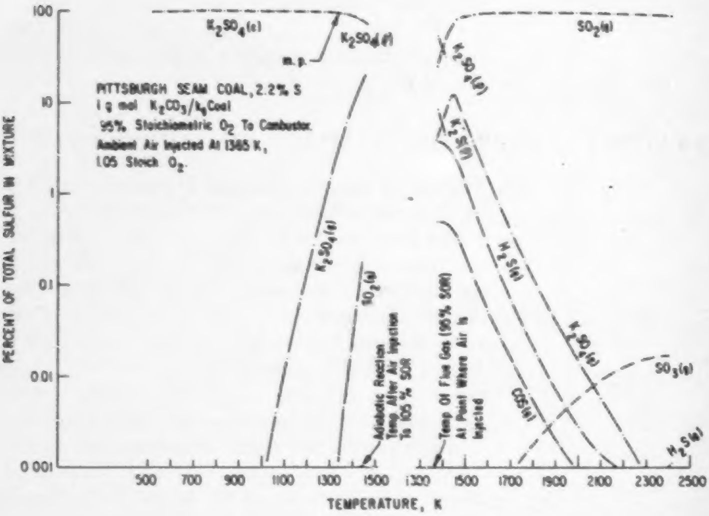


Fig. 5.13—Distribution of sulfur compounds in two-stage combustion

combustor and ambient air admitted at combustion gas temperature of 1365 K (see Fig. 5.13). Even at 95% stoichiometric air, the most abundant sulfur compound at elevated temperatures is gaseous SO_2. Gaseous K_2SO_4 exceeds H_2S and COS. Gaseous H_2S accounts, at most, for only 4% of the available sulfur. As the temperature drops, the condensed liquid phases of K_2SO_4 and K_2S are formed. With injection of ambient air to increase oxygen to 1.05 of stoichiometric, the temperature is raised by 75 K. As the gas is further cooled, virtually all of the sulfur is incorporated in condensed K_2SO_4. Thus, complete removal of SO_2 is not hampered by two-stage combustion. As an additional precaution against corrosion, it might be advantageous to add air above 1400 K, the temperature at which liquid K_2S is formed. If gaseous H_2S is a problem in corrosion, it may be preferable to allow a somewhat higher NO_x emission by adding the ambient air at 1500 K, at which temperature gaseous H_2S accounts for only 1.2% of the total sulfur compounds.

5.7 SUMMARY

MHD power stations have the potential for substantially decreasing pollutant emissions and meeting environmental clean air and thermal pollution standards.

Existing technology is capable of achieving adequate removal of particulate impurities from flue gases.

A high degree of combustion products purification by removal of sulfur compounds is achieved through interaction with the seed material that is injected into the combustion products to increase electrical conductivity. At a seed level of 1.5 g-mol of K_2O/kg of coal, practically all the sulfur oxides produced in the burning of fuels containing 4% sulfur can be removed. Regeneration of the captured seed material yields elemental sulfur. The regenerated seed can then be returned to the MHD cycle.

Two-stage combustion with a fuel-rich mixture in the first stage and with staged cooling at a predetermined rate downstream of the MHD channel lowers the initially high concentration of NO_x in the flue gases to 200 ppm, a significantly lower concentration than that of a conventional coal-fired power station. Successful pilot-plant experiments have been carried out with the combustor operated with 90 and 95% of the stoichiometric air requirements. Secondary air injected for afterburning in the zone below 1600 K does not lead to NO_x formation.

Optimum conditions for NO_x control depend upon trade-offs between a number of conflicting requirements and will vary between installations because of differences in the type of fuel used and in the practical constraints on a particular system design. The questions to be addressed are the definition of the optimum conditions for the fuel equivalence ratio in the combustor, the temperature at which the secondary air should be injected, and the cooling rate downstream of the diffuser.

Reducing the air supply to the combustor provides the potential for lower NO_x emissions, but has the disadvantages of increased residence time for gasification/combustion of solid fuels and greater increase in temperature during secondary air injection. The temperature at which secondary air is injected should be low to prevent excursions above 1800 K, where NO_x formation reactions become significant; but reducing the temperature at which the air is added increases the volume of the superheater-reheater and the difficulty of burning the CO. A low cooling rate (200 K/s or less) downstream of the diffuser is desirable to minimize NO_x emissions, but is associated with the penalty of a large volume for the radiant boiler. Higher cooling rates may be possible in the presence of heterogeneous reactions provided by either the ash and any char present in coal combustion products or the use of alumina heat exchangers; optimization of cooling rates under such conditions requires further research.

Technical-economic analysis indicates that electricity from an MHD power station can be made 10-15% less expensive, through production of nitric acid from NO_x taken from the stack.

Thermal pollution from an MHD station ($\eta = 50\%$) is less by a factor of 1.5 than that of a conventional power station ($\eta = 40\%$), and by a factor of 2.0 than that of an atomic power plant ($\eta = 33\%$). This advantage results from higher efficiency and the fact that the steam turbines comprise a smaller portion of the capacity.

Nomenclature

Symbols

g	gas
K_{NO_2}	measured concentration of NO_2 in atmospheric air
$K_{NO_2}st$	maximum allowable concentration of NO_2 in atmospheric air
K_{SO_2}	measured concentration of SO_2 in atmospheric air

$K_{SO_2}st$	maximum allowable concentration of SO_2 in atmospheric air
l	liquid
Q_{total}	total heat rejected
α	stoichiometry
η	overall thermal efficiency

Acronyms

EPA	Environmental Protection Agency
PERC	Pittsburgh Energy Research Center

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